

Chapter 6

Power Generation from Biomass and Waste

The thermal use of biomass or waste is a cheaper and more technically complete option for reducing CO₂ emissions compared to other renewable energy sources. Possible biogenous fuels include forestry and agriculture residues such as wood and straw, and also specially cultivated reproducible raw materials such as *Miscanthus sinensis*, whole cereal plants, poplars or willows.

For the conversion of biomass into electric power and heat, a great number of processes are available. They can be classified according to the following:

- Fuel type: biomasses (such as wood or straw), organic residues, municipal solid waste, sewage sludge or refuse-derived fuel
- Fuel conversion: combustion or gasification (atmospheric or pressurised)
- Power production: combustion engine, gas turbine, steam turbine, Stirling motor or fuel cell
- Output capacity
- Products: electricity or heat or both combined
- Exploitation of biomass alone or combined with fossil fuels

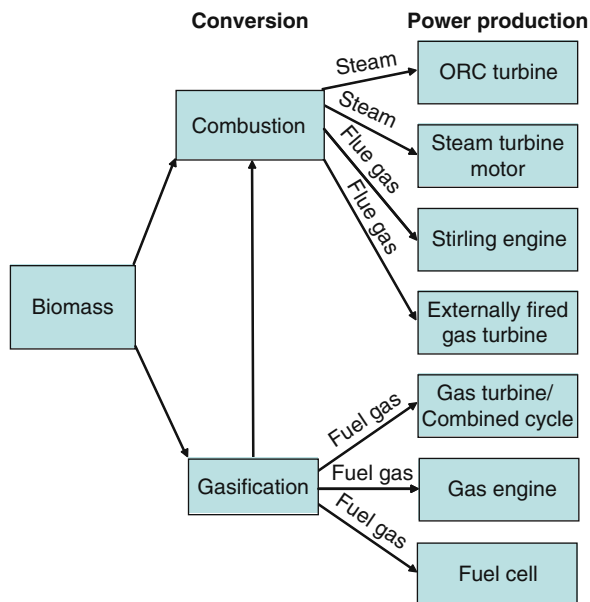
6.1 Power Production Pathways

The pathways for heat or electric power production from solid biomass and waste can be subdivided into systems with combustion and systems with gasification. A systematic arrangement of the individual steps in each pathway is presented schematically in Fig. 6.1.

6.1.1 Techniques Involving Combustion

In combustion, the chemical energy of the fuel is converted into sensible heat which the flue gases carry, which is then transferred to a working medium in a heat exchanger. The working medium expands in a turbine or other such device to create mechanical energy for power production.

Fig. 6.1 Pathways for the production of power from biomass



Steam Turbines and Steam Engines: Production of power by steam turbine is the widest spread electrical power generation process in thermal power plants. The efficiency of the power generation is determined by the live steam conditions in front of the turbine and the exhaust steam conditions. For economic reasons, lower plant outputs mean lower achievable efficiencies. In the range of several megawatts, it is possible to achieve around a 20% plant efficiency and in the range of several tens of megawatts, about 30%. In comparison to this, the maximum efficiencies reach about 45% in the capacity range of several 100 MW. Steam turbines are on the market for capacities starting from 1 MW; for capacities less than 1 MW, steam engines are used. In the low capacity range, the high costs of the steam power process, which uses turbines, are a disadvantage.

Organic Rankine Cycle (ORC) Processes: The ORC process is based on the Rankine process, just as the conventional steam process is. Instead of water, however, the working fluid is of organic origin and has low boiling and condensation temperatures. Accordingly, the ORC process is suited to heat exploitation at a lower temperature. Due to the low temperature of the working fluid, the resulting efficiencies for biomass applications from a thermodynamic point of view must be lower than in a steam process, because the maximum steam temperature of the organic medium is in the range of 250°C. Electrical efficiencies for ORC plants in operation are in the range of up to 15%; however, it has to be considered that nearly all are operated as CHP plants (ADMONT 2001; Obernberger 2003). For the purposes of comparison, this would correspond to an electrical efficiency of about 20–24% for a pure power cycle. Improvements in the ORC power production process are currently

being introduced by employing two parallel flows of the organic medium to improve the heat transfer (i.e. the split process) (Turboden 2009).

Externally Fired Combined Cycle (EFCC) Processes: This closed gas turbine process works with a high-temperature heat exchanger which can be operated using the exhaust gas from the firing. Here, the energy is transmitted to pre-compressed air which is then expanded in a turbine. Metallic materials may be used for the construction of smaller capacity gas turbines because of the lower turbine entry temperatures. Therefore, the EFCC process is ideally suited to new and existing biomass-fired furnaces for decentralised electric power and heat generation. The problems arising in gas cleaning in biomass gasification are avoided. In general, an efficiency in the order of magnitude of 25% seems to be achievable (Kaltschmitt et al. 2009).

Stirling Engines: The Stirling engine is a type of expansion engine. In contrast to combustion engines, the piston is moved not through the expansion of exhaust gases from internal combustion but through the expansion of a constant, closed gas volume, caused by heat transfer from an external heat source. This way, the generation of power is separate from the heat source, which means such engines can be fuelled by heat from biomass combustion plants. In the case of biomass-fired furnaces, there is a conflict concerning the design. On the one hand, the aim is to heat the smallest possible gas volume to high temperatures, on the other hand, for a high-efficiency and a low-heat flux, the heat exchanger should be designed at the largest possible scale. For natural gas-fired plants with capacities of between 10 and 40 kW, it is possible to achieve efficiencies of 20–25% by utilising the hot waste air as combustion air. Since this variant cannot be used in biomass-fuelled installations because of slagging, the efficiencies for biomass-fired engines range between 10 and 15%.

6.1.2 Techniques Involving Gasification

In gasification, the fuel is converted at air-deficient conditions into combustible gas by an addition of air or another gasification agent (such as water vapour). The combustible gases that are produced can be fed to a machine driven by internal combustion. Particular consideration is given to the gas engine and the gas turbine, each of which involve different efficiencies, costs and gas quality requirements.

Gas Engines: In gas engine applications, the product gas is mixed with air, conducted to the engine and burned in its cylinder according to either the spark ignition or the diesel principle. The mechanical work of the engine is converted into electrical energy by means of a coupled generator. Engines in conjunction with atmospheric fixed bed or fluidised bed gasifiers are suitable for capacities from about 50 kW_{el} to 10 MW_{el}. With gasification and subsequent motor-driven use of the combustible gas, it is possible even at small capacities to achieve high electrical efficiencies above those of combustion-based processes. For capacities around 500 kW_{el}, net efficiencies between 24 and 31% are indicated (Vogel 2007).

Gas Turbines: For outputs of 5 MW_{el} and over, it is reasonable to use gas turbines. Suitable gasifiers in this range are atmospheric or pressurised fluidised beds. Gas turbines have the potential of increasing the efficiency up to about 45% by installing a downstream heat recovery boiler with a steam turbine (output 20 MW_{el}).

Besides that, it is possible to conceive using gas turbines in innovative technologies presently still in development, such as fuel cells. In fuel cells (FCs), the chemically bound energy of the fuel is directly converted into electrical energy, with higher efficiencies than conventional technologies can achieve. Coupling of single FC technologies with downstream gas and steam turbine processes presents itself as a means for better fuel exploitation. FC technology, however, is still in different stages of development, depending on the FC type.

6.2 Biomass Combustion Systems

Biomass combustion systems can be classified by the same principle as used for coals (see Sect. 5.1) – that is, into fixed bed, fluidised bed and pulverised fuel systems. A lot of technologies have been derived from coal combustion systems; in the smaller capacity range, especially, a large variety exists (Nussbaumer 2003; Van Loo and Koppejan 2008; Eltrop et al. 2007; Kaltschmitt et al. 2009). In the course of this chapter only the most widespread systems are presented.

6.2.1 Capacities and Types

One essential criterion for the choice of the combustion system is the size of the plant that shall be built. In this respect, the classification distinguishes between small furnaces of up to a 15 kW thermal capacity, medium-scale plants of up to 1 MW and large-scale plants (Kaltschmitt 2001):

- Small furnaces are used in the household sector for hot water and room heating, with capacities of up to 15 kW_{th}. They will not be considered further in this text.
- Plants up to a thermal output of 1 MW_{th} are used in commerce and trade. The firing systems widely employed are shaft and underfeed firing furnaces. Investigations into the emission behaviour of plants in service in industry and trade show that, during changes of operating mode (i.e. start-up, shutdown, load change), higher emissions of dust, carbon monoxide and hydrocarbons are produced through incomplete combustion. These emissions are often caused not only by intermittent fuel feeding practices at start-up and shutdown but also from operation at part load, when the output is controlled by connecting and disconnecting the fuel supply and the fuel/air ratio is not set optimally. Newly developed underfeed firing systems for woodchips show that combustion and operation at low emission levels are possible even in this output range of up to 1 MW_{th}.
- Plants with capacities higher than 1 MW_{th} which are used for the production of heat, process steam and electricity are usually operated as combined heat and

power production plants (CHP plants). The upper capacity limit of plants fired exclusively with biomass is seen today as 50–100 MW_{th}, because transport and the logistics of the fuel supply become too complex and costly at higher capacities. In this capacity range, the predominant firing types are stoker-fired furnaces, which are suited to fuels that are moist, problematic and/or in lumps, and require little in terms of fuel preparation. Fluidised bed furnaces are attractive because of their lower emissions but have more complex constructions and are therefore economical only for output capacities above 10 MW_{th}. Furnaces designed for pulverised fuel are suitable for firing biomass, in particular if the fuel is available in a small-particle form. Pulverised fuel-fired furnaces are the dominating combustion technology for coal in large plants because of their high power density, good controllability and complete combustion. For fuels with fine and coarse fractions, it may also be practical to combine pulverised fuel and stoker firing. In Denmark, the so-called cigar burner proved to be a reliable technology for the combustion of straw bales.

- Co-combustion: Besides using biomass as the only feedstock, it is also possible to fire it combined with other, preferably solid, fuels. The respective firing and flue gas cleaning systems must be tested for their suitability in this case.

6.2.2 Impact of Load and Forms of Delivery of the Fuel Types

Besides considering the power plant capacity, the choice of the firing system has to take into account the shape (shavings, chaff, pellets, bales, etc.) in which the biomass is available. Figure 6.2 presents the field of application of firing systems as a function of the plant size and the biomass shape.

- Pieces of firewood (logs) can be used in shaft or pusher-type grate furnaces, depending on the thermal input.

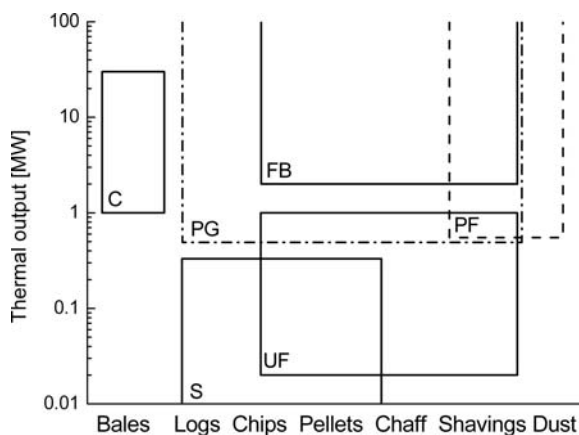


Fig. 6.2 Combustion systems as functions of plant size and biomass shape (PF pulverised fuel, S shaft furnace, UF underfeed firing, PG pusher-type grate, FB fluidised bed furnace, C cigar burner)

- Woodchips can be used in almost all firing systems (i.e. shaft, underfeed, stoker-fired and fluidised bed furnaces).
- Preparation into the form of woodchips is the standard technology for woody biomass types.
- Shavings of wood can be burned in underfeed firing up to a determined fraction. In pusher-type grate, pulverised fuel and fluidised bed firing systems, there are no such restrictions.
- Powdery biomass can be used only in pulverised fuel firing.
- Straw bales can be directly burned in cigar burners, which are used at capacities over 3 MW.
- Chaff from herbaceous biomass can be used in underfeed, in pusher-type and, to limited extent, in fluidised bed firing systems.
- Pellets are well suited for use in firing systems, with advantages similar to chopped material.

For herbaceous biomass types, there are essentially two competing methods of delivery. In the first, the biomass is mown in the field and pressed into bales. Bales are either fed as a whole to the cigar burner or are undone and cut into chaff before being fed to the combustion plant. The second delivery method involves the herbaceous biomass being pelletised in the field, thus increasing the density for transport. For underfeed, grate and fluidised bed firing systems, the fuel is delivered ready in the form most suitable for combustion.

In Denmark, the delivery method that has become generally accepted for the thermal utilisation of straw is the pressing into large-size Hesston bales, which can be performed at a relatively low cost using state-of-the-art technology (Kaltschmitt 2001; Nikolaisen 1992).

In biomass combustion, a high fuel storage capacity is required because of its low energy density. Storing on-site at the combustion facility over a long period is only possible for smaller facilities, so for plants of the megawatt order of magnitude or greater, biomass is usually stored on the producer's premises and delivered daily to the combustion facility. Often, these plants have an on-site fuel storage capacity of just 2–3 days fuel consumption.

6.2.3 Furnace Types

6.2.3.1 Shaft Furnaces

In the lower capacity range, from 20 kW up to about 250 kW, shaft furnaces are used for the combustion of both lump wood residues and woodchips. The fuel types that can be used in these plants usually do not require additional preparation (Eltrop et al. 2007; Spliethoff and Hein 1995a). Figure 6.3 shows an example of a shaft furnace with lateral burnout for thermal capacities between 50 and 150 kW.

The furnace design is based on the principle of bottom, or lateral burnout, which means that the flame extends from the side or the bottom into the combustion cham-

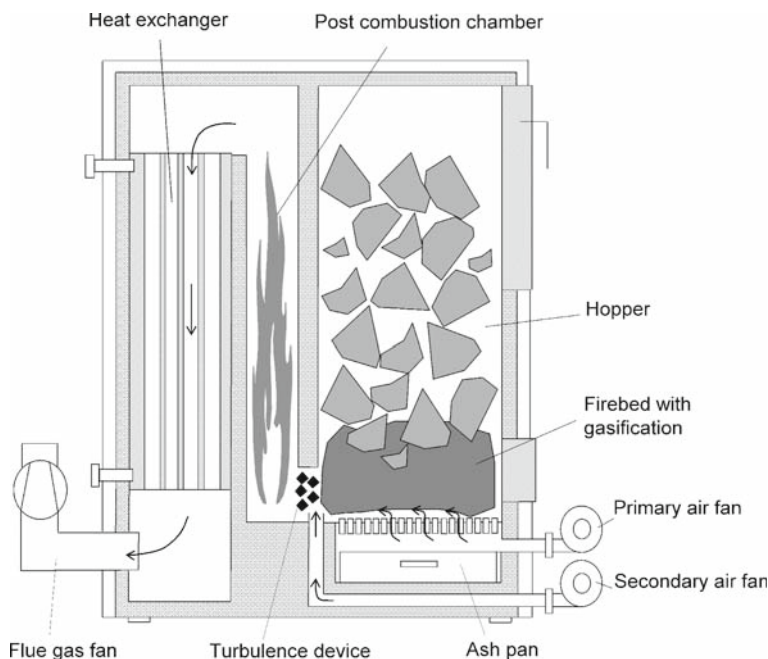


Fig. 6.3 A shaft furnace with lateral burnout (Kaltschmitt 2001)

ber. The air, supplied by natural draught or by fan, is divided into primary and secondary air in modern installations. The primary air is injected below and at the side of the grate, while the secondary air is directed into the combustion chamber. As a rule, these furnaces are fed manually and operated intermittently. The degree of automation and potential for control are smaller than for plants with continuous feeding. After the frequency and quantity of feeding, further control is generally limited to combustion air throttling. Modern firewood boilers with two-stage control allow adjustment of the firing to produce low emissions. The output in this case is controlled by the primary air flow; the secondary air, controlled by the exhaust gas composition, ensures the complete burnout.

The lack of the need for fuel preparation, the simple firing technology and the relatively low cost of purchase are the reasons for the widespread use of this firing type in the given capacity range.

6.2.3.2 Underfeed Firing

Underfeed firing systems, which are available on the market in a broad capacity range – from $20 \text{ kW}_{\text{th}}$ to 6 MW_{th} – are suitable for the firing of chips, pellets, shavings and, to a certain extent, for pulverised wood residues as well – those with a moisture content between 5 and 40%. The maximum size of the fuel is limited by

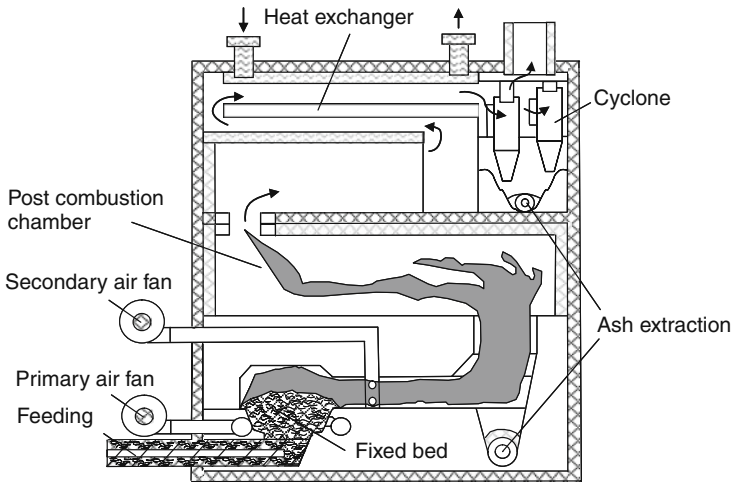


Fig. 6.4 Underfeed firing (Kaltschmitt et al. 2009)

the fuel feeding system. The fuel should be low in ash, finely grained and of a homogeneous structure.

The firing systems (Fig. 6.4) are automatically fed by screw conveyors directly from a silo. In a great number of the furnaces, the fuel is transported from below into a trough inside the combustion chamber, a so-called retort, then dried and devolatilised.

The pyrolysis gases and primary air enter the glowing bed layer from below, ignite and burn to completion following the addition of burnout air. Underfed fires are as a rule self-igniting and in principle controllable to an acceptable degree by co-ordinated fuel and air supply. Continuously working screw feeders and the small amounts of embers and fuel in the furnace result in operation with little emission of pollutants.

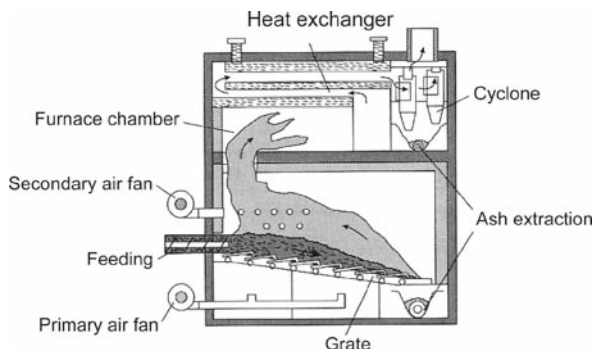
This combustion technology is widely employed for the thermal utilisation of residues from wood processing, because it

- works almost fully automatically,
- is built using simple technology and fewer components in comparison to other firing types such as pulverised fuel or stoker firing systems and
- is economically attractive, even if the storage, feeding and potential need for fuel preparation equipment is taken into account (Van Loo and Koppejan 2008; Spliethoff 2000).

6.2.3.3 Stokers

For capacities of around 1 MW_{th} and higher, stoker-type furnaces are the dominant technology for the combustion of biomass. Figure 6.5 shows a forward pusher-grate furnace, which is the dominant system used for woody biomass. Other grate

Fig. 6.5 A forward pusher-grate furnace (Kaltschmitt et al. 2009)



firing technologies, such as travelling or reverse reciprocating grates, can also be used; they are described in Sect. 5.5. Problematic fuels, for example, moist wood residues or high-ash bark residues can also be fired in these systems. In stoker firing, it is also possible to burn straw as the only fuel – as the practice in Denmark demonstrates.

The grate is either fed from the fuel storage via a hopper, by screw conveyor or hydraulic conveyor. The feedstock moves from the feeding point to the grate end at the grate speed. The fuel dries, pyrolyses and burns completely as it is pushed down the grate.

Primary air enters from below the grate and through the grate bars, and secondary air is injected above the grate and ahead of the partly firebrick-lined secondary combustion zones. In stokers, low fuel qualities can be combusted effectively because the residence times of the fuel and the combustion air flow can be adjusted to a wide range of fuel properties. Compared to underfeed-type furnaces, moving-grate stokers are harder to control and less suitable for fast load changes due to the greater fuel loads inside the furnace.

With their complicated plant mechanics, stokers are economical only from capacities of around 1 MW_{th} and above. For the combustion of wood and wood residues, these furnaces are partly operated in connection with a direct firing system for pulverised residues.

Stokers are also suited to the combustion of herbaceous biomasses such as straw, *Miscanthus* or grass. However, the ash fusion temperatures, lower than for wood, may result in fuel caking, which impedes the passage of air and thus combustion. In these cases, temperatures below the ash softening point can be set by lower bulk heights and grate heat release rates. Grate raking is not a reliable remedial action if it risks causing incomplete burnout (Biollaz and Nussbaumer 1996).

6.2.3.4 “Cigar Burner” for Herbaceous Biomass Bales

The so-called cigar burner of the Volund Company is in principle a grate firing system, as can be seen in Fig. 6.6. The bales, however, are not deposited on the grate in exactly the delivered form but first ignited at the front before being gradually pushed into the combustion chamber.

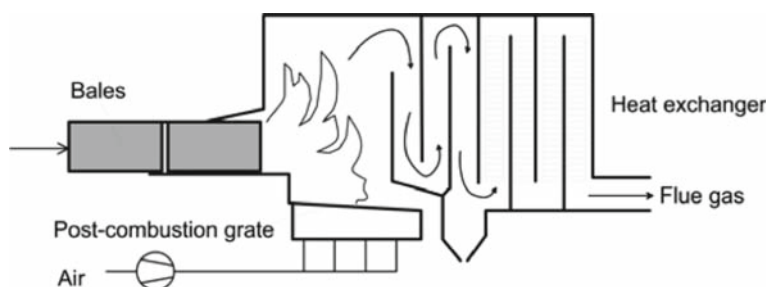


Fig. 6.6 A cigar burner

Unburned straw layers that break off fall into pieces onto the grate and burn to completion while moving along it. The ash gets discharged as a result of the forward movement of the grate. Water cooling and flashback dampers inside the bale charger prevent a burning back of the bales.

The advantages of this technology are the relatively minor preparation needed for the fuel, the continuous fuel supply and the relatively simple construction of the plant. The disadvantages are the narrow range of fuels and the restriction to one bale type. Burning straw layers falling onto the grate at intervals may cause increased CO emissions.

The bale shape and the minimum feed rate determine the minimum stable capacity of this firing type. Because of the complex automatic control technology of the storage and feeding installations, this firing type is justifiable from an economical point of view only for capacities of about 2–3 MW_{th} and above. A modified version of this type for smaller outputs involves a preparation step where the large bales are cut into several slices to be put onto the grate.

In Denmark, several plants of this type are in service, predominantly for straw bales, at capacities from 3 to more than 20 MW_{th}. The boilers can be used both for heat and for electricity production (Van Loo and Koppejan 2008; Nikolaisen 1992).

6.2.3.5 Fluidised Bed Combustion (FBC)

In fluidised bed combustion, the prepared fuel is burned at 800–900°C in a fluidised bed consisting of 95–98% inert material and only 2–5% combustible material. The process-related intensive mixing and combustion, the excellent heat transfer in the fluidised bed and the decoupling of the particle residence time and the flue gas velocity allow a wide range of fuels to be used. In particular, a wide range of moisture contents, compositions and fuel preparations can be exploited. Fluidised bed furnaces are particularly suited to the combustion of several, even very different, fuel types. In addition, fluidised bed combustion has advantages over grate combustion when sludge fuels are used.

Given the low combustion temperature, problems such as slagging and fouling are less severe in FBC furnaces. There is a risk of fluid bed sintering, but only when firing very high alkali fuels, such as straw. This risk may rule out the application of fluidised bed technology for such fuels if they are to be used in monofuel combustion. The characteristics of bubbling and circulating fluidised bed combustion are described in detail in Sect. 5.4.

Today, the design of modern, stationary fluidised bed firing systems for biomass clearly deviate from the design of a classic stationary fluidised bed for firing coal. When firing biomass, there is no need for an in-bed heat transfer surface because the bed is operated with a high deficiency of air and so only part of the fuel heat is released in the bed. By setting the air ratio in the fluidised bed, it is possible to reliably control bed temperatures between 650 and 800°C. The air ratio depends on the heating value of the fuel and the temperature required to assure the prevention of sintering of the bed. For wood chips the air ratio is typically in the range of 0.35–0.45 and the temperature is below 800°C (Bolhar-Nordenkamp et al. 2006). The temperature at which sintering of the bed can occur depends mainly on the ash composition of the fuel. When biomasses with high alkali contents are used, such as straw, sintering temperatures significantly below 800°C develop. The addition of the remaining air is performed in several stages in the freeboard of the furnace. In this process, for the sake of cleanliness, temperatures of about 1,050°C for clean wood should not be exceeded when air is added so that fouling in the freeboard is prevented. The temperatures can be controlled by adding the secondary air in several stages and by heat dissipation. The air-staged mode of operation leads to low emissions of NO_x . The principle of air-staged fluidised bed combustion is shown schematically in Fig. 6.7 and compared to the design of a stationary fluidised bed for coal.

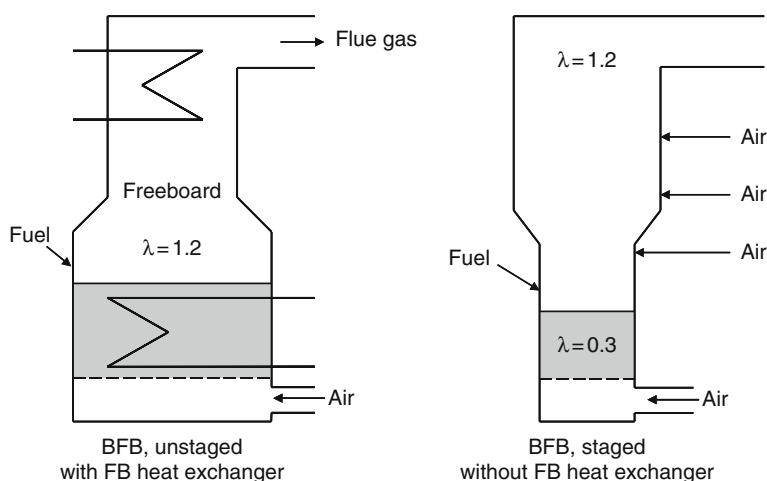


Fig. 6.7 Staged BFB combustion (biomass) in comparison to unstaged BFB combustion (coal)

The fuel preparation for a bubbling FBC furnace is a similar process to that used in a stoker-fired furnace. In both plants, wood cut into pieces smaller than 90 mm can be used. A circulating FBC furnace requires pieces smaller than 30–50 mm. Since fluidised bed combustion involves high capital costs, it can be operated economically only in larger units (bubbling FB greater than 10 MW_{th}, circulating FB greater than 30 MW_{th}). At Stadtwerke, Leipzig, a CFB with a capacity of 55 MW fuel input (clean wood) went into operation in 2005. The plant features an electrical efficiency of 37%, which is achieved with high steam conditions of 535°C, 130 bar and reheating (to 535°C) (Beckert and Schaarrschuch 2007). Such advanced conditions are the exception for a plant of this size.

Fluidised bed combustion, as a novel technology, competes with the predominant biomass combustion technology, i.e. the stoker. For monofuel combustion of wood, FBC is a proven technology, but not yet for herbaceous biomass. In the Scandinavian countries and increasingly in other countries, too, wood residues and sludge from paper and pulp production are mostly fired in fluidised bed furnaces that are an integrated part of the production process. Due to the economical output capacity, especially of circulating fluidised bed furnaces, biomass is used along with other fuels such as peat, coal and waste in such plants (Gockner and Rechberger 2008; Bolhar-Nordenkamp et al. 2006; McCann and Simons 1997; Seemann et al. 2008).

6.2.3.6 Pulverised Fuel Combustion (PFC)

The pulverised fuel furnace is a practical solution for biomass, in particular when the available fuel is already very fine. This is often the case in wood-processing enterprises, which usually employ high-speed machines, resulting in fine residues. For fuels with a certain fraction of coarser materials, a PFC furnace with a burnout grate may be reasonable. PF-fired furnaces stand out because of their high power densities, high furnace efficiencies and good controllability. They are offered on the market at capacities ranging between 500 kW_{th} and 50 MW_{th}.

Pulverised fuel furnaces for biomass combustion are usually designed as direct firing systems, i.e. by cyclone or through a muffle.

Figure 6.8 shows the principle of direct firing through a muffle for the combustion of pulverised wood. The fuel is injected tangentially into the combustion chamber together with the primary air. A whirling flow, similar to a vortex flow, develops. Larger particles are blown into and deposit in the rear of the combustion chamber, burning out after they have landed. Fine particles burn out while still airborne. Secondary air is injected at a constriction in the muffle. This constriction at the end of the muffle causes a rise in the velocity and thus a good mixing of the flue gases with the secondary air. Modern and bigger cyclone suspension furnaces are operated with continuous control – the fuel feeding is adjusted to the firing rate required at a given time and the combustion air supply is adjusted exactly to the fuel charge. These firing systems help to keep the emissions of unburned flue gas components

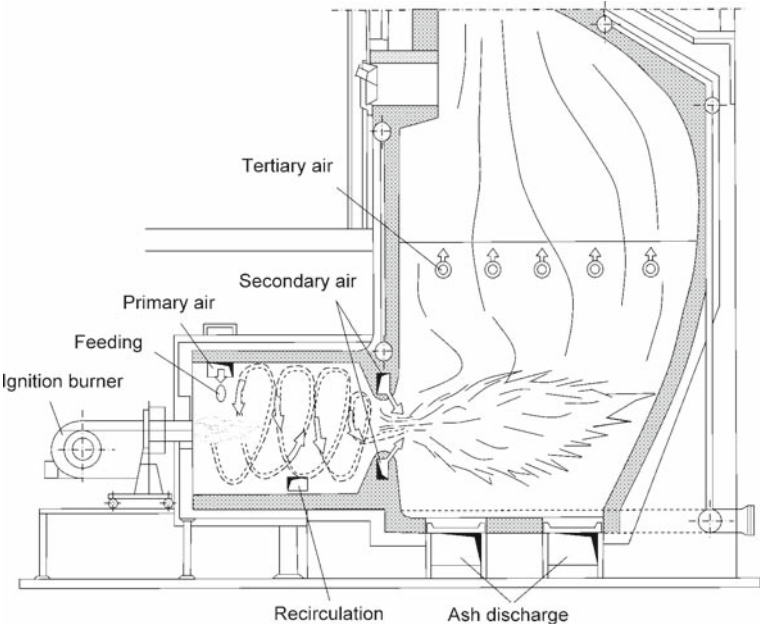


Fig. 6.8 A pulverised fuel muffle furnace (Kaltschmitt et al. 2009)

to a low level. The exact matching of fuel and combustion air makes it possible to apply effective primary measures of NO_x reduction.

6.2.4 Flue Gas Cleaning and Ash Disposal

The application of emission reduction techniques depends on the plant size, the fuel to be used and the emission limits to be complied with. Table 6.1 shows typical emission concentrations for the untreated gas (before flue gas cleaning) and after dust

Table 6.1 Typical flue gas emissions of woodchip combustion plants (Spliethoff 2000)

Emissions (mg/Nm ³)	Typical value	Range	Emission limit, Germany (TA Luft) ^a (mg/Nm ³)
SO ₂	170	50–350	–
NO _x as NO ₂	250	100–400	250
Particle (raw gas)	500	200–800	20–100 ^b
Particle, multicyclone	300	200–400	
Particle, flue gas condensation	40	20–50	

^a 1–50 MW_{th}

^b Depending on capacity

collectors in wood chip combustion plants. These values, in practice, can strongly deviate beyond the given ranges in the table. They are compared to the emission limits of the *TA Luft* (the German Clean Air Code) for wood-firing appliances. It is obvious that, for larger biomass combustion plants, the design has to include flue gas particulate collectors. In biomass-fired plants, aside from dust collection, further flue gas cleaning units are not commonly installed.

6.2.4.1 Particulate Control

Dust collection for herbaceous fuels is more demanding than for woody biomass. Due to the greater lump sizes and the higher density of the fuel, the ash produced in wood combustion is coarser. With herbaceous biomass, the higher alkali, chlorine and sulphur contents trigger the development of salts (KCl , K_2SO_4), which form ultrafine fly ash particles that can be separated effectively only by a particulate filter such as an electrostatic precipitator (ESP) or a fabric filter.

Multicyclones used in wood combustion efficiently only remove particles larger than $10\text{ }\mu\text{m}$ with tolerable pressure losses, which, in chip and bark combustion, results in residual flue gas dust contents of 120 to around 400 mg/Nm^3 . The emission limits, dependent on the plant size, cannot be met in most instances, which means another dust collecting unit has to be installed downstream. For woody biomass, common practice is to use cyclones for coarse separation and ESPs or fabric filters for fine separation. This way it is possible to reduce dust contents to between 10 and 50 mg/Nm^3 . ESPs are the preferred technology. For small plants the higher pressure loss of fabric filters exclude their application; for plants larger than 10 MW, ESPs are more economical (Kaltschmitt et al. 2009; Hasler and Nussbaumer 1996; Johnsen and Svendsen 1997; Obernberger 1997; Biollaz and Nussbaumer 1996).

ESPs can be problematic if used when firing herbaceous biomass, because the dust of dry straw ash has a higher resistance than coal ash. However, if ESPs are employed at lower flue gas temperatures, the fly ash will absorb water and the resistivity will decline. If temperatures become too low, the ash becomes sticky and will cause fouling problems. The result is a narrow temperature window around 115°C for the use of an ESP (Johnsen and Svendsen 1997). The separators well suited to these applications seem to be fabric filters, which, besides their high collection efficiency, have the advantage of a higher degree of capture of chlorine and sulphur in the fly ash, which is favoured by the intensive contact between flue gas and ash on the surface of the filter bags (Obernberger 1996).

For all the fuels mentioned, wet cleaning processes, in which the flue gas is conducted through a scrubber where spray water is injected, are also used. During this process, the dust particles are bound and discharged with the water. In addition, acidic pollutant gases, such as SO_2 and HCl , can be removed by wet cleaning. However, the wastewater creates an additional disposal problem (Nikolaisen 1992).

When moist fuels are used, flue gas condensation plants for heat recovery can double as particle collectors, if needed in combination with wet dust removal.

The flue gases are cleaned in a multicyclone and afterwards the water vapour is condensed for heat recovery. This way, it is possible to achieve dust contents of 40 mg/Nm^3 . This method is applied only for very moist biomasses, having more than 30% moisture content – usually woodchips (Eltrop et al. 2007).

6.2.4.2 Nitrogen Oxides and Sulphur Oxide

The combustion of wood or straw does not require downstream DeNO_x or desulphurisation installations, such as are state of the art in pulverised coal-fired power plants.

The sulphur contained in the fuel, though mostly released in a gaseous form during combustion, can be captured in the fly ash. Downstream of chip or bark combustion furnaces, for instance, the ash removed by multicyclone captures 40–70%, while during fine dust collection in a fabric filter, between 60 and 90% of the sulphur is captured. Only the remaining gaseous portion is released as SO_2 emissions. With straw or whole plants, the capture rates of the respective ashes are about 45–50%. The capture in the ash depends on the concentration of alkalis and alkaline earths (especially Ca), on the combustion temperature and the dust collection method employed. Due to the low sulphur contents of straw and wood, it is possible to do without desulphurisation in systems firing these fuels (Oberberger 1996).

The low fuel nitrogen content of wood results in only minor NO_x emissions during combustion. For fuels with higher fuel nitrogen contents, such as straw or whole plants, it is possible to apply primary measures to meet the emission limits. Primary measures are effective because of the volatile matter fraction of the fuel nitrogen. Figure 6.9 shows NO_x emissions measured in stokers in service (Nussbaumer 2003; Biollaz and Nussbaumer 1996).

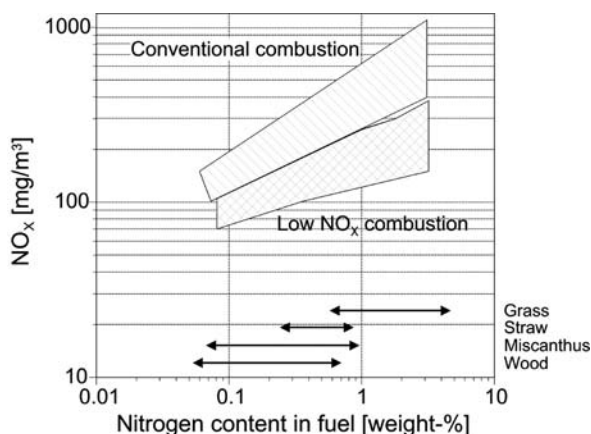


Fig. 6.9 NO_x emissions from biomass-fired stokers (Biollaz and Nussbaumer 1996)

6.2.4.3 Chlorine

Chlorine can be captured in fly ash as well. At woodchip and bark combustion facilities, capture rates between 40 and 80% have been measured, with the rates for straw and whole plants ranging around 80–85% when applying a downstream fabric filter. Even with an 85% capture of chlorine, the German limit of 30 mg/m³ for straw- and whole-plant firing systems is exceeded when there are fuel Cl concentrations over 0.15% by weight in the dry matter.

Higher chlorine concentrations in consequence require secondary measures such as a scrubber or ultrafine dust collector combined with dry sorption (Biollaz and Nussbaumer 1996). If fabric filters or ESPs are used, the flue gas is mixed with an alkaline solid sorbent such as calcium hydroxide (Ca(OH)₂) before entering the filter. In wet dust collection processes, the removal of acidic pollutant gases without additives is possible but is more effective when alkaline additives are used (Hasler and Nussbaumer 1996).

High chlorine contents may trigger dioxin formation. High combustion temperatures and a fast cooling down of the flue gas to the low-temperature area are a means to ensure low dioxin emissions. Measurements taken at various biomass-fired plants in service for different fuels (woodchips, bark, straw, whole plants) show that the applied firing technologies – underfeed, grate and cigar burner – in normal operation comply with the emission standards laid down in the 17th *BlmSchV* (Ordinance on Incinerators for Waste and Similar Combustible Materials) (Obernberger 1996).

6.2.4.4 Ash Utilisation

The residual matter from the combustion of natural wood can be used as a fertiliser in agriculture and forestry. While the dioxin and furan contents of these ashes are ecologically harmless, their heavy metal content must be assessed. The enrichment of heavy metals in the fine particulates of wood ash prohibits the utilisation in forests. Certain highly volatile heavy metals such as cadmium, lead and zinc concentrate in the finest ash fraction in the ESP or the fabric filter. Waste disposal strategies aim at enriching the heavy metals in the ultrafine filter ash (which is disposed of after removal) and utilising only the coarse ash from the cyclone together with the grate ash (Kaltschmitt et al. 2009; Obernberger and Biedermann 1996).

In Austria, the fertilisation of forests, fields and grassland with cyclone and grate ash from wood combustion is a widespread practice, controlled by the authorities. Filter ash and sludge from flue gas condensation are always sent to landfill (Kaltschmitt et al. 2009). In Denmark, woodchip ash is often used to fertilise the forest area from which the wood was harvested. The ash from straw-fired thermal power plants is mainly used as a fertiliser in agriculture (Nikolaisen 1992). In Germany, the utilisation of biomass ash as a fertiliser and soil conditioner in agriculture and forestry falls under the regulation for fertilisers. Furnace ash from monofuel combustion of untreated biomass can be used for the production of fertilisers and soil conditioners, whereas cyclone or ultrafine fly ashes cannot be used as a matter of principle. When using herbaceous biomass ash as a fertiliser or when

using any ash on agricultural fields, however, the heavy metal concentration limits of the Biowaste Ordinance have to be complied with. These limits are significantly stricter than the limits of the sewage sludge regulation (Eltrop et al. 2007).

The combustion of waste wood, compared to natural wood, involves increasingly stringent requirements for flue gas cleaning, for instance, for HCl, which may require a secondary removal stage. With scrap or residual wood from chipboard fabrication, it may additionally be necessary to remove nitrogen. In waste wood combustion, heavy metals are removed for the most part together with the fly ash – which restricts the utilisation of the ash. Mercury, however, which can be emitted with the flue gas, is found only in small concentrations in waste wood and is therefore usually not a problem (Hasler and Nussbaumer 1996).

6.2.5 Operational Problems

The chemical components most critical for the smooth operation of biomass combustion plants are the chlorine and alkalis present in the ash. Causes and effects of slagging and corrosion are discussed in Sect. 5.10. Deposits of alkali chlorides are the reason for chlorine-induced corrosion, which limits the application of higher steam temperatures. The corrosion mechanism is similar to waste incineration (see Sect. 6.4). While there is little corrosion when woody biomass is the fuel, due to the low chlorine and alkali concentrations, serious corrosion problems may occur with herbaceous and petiolate biomass (straw, *Miscanthus* and whole plants).

As steam and tube wall temperatures rise, the corrosion rate increases. Figure 6.10 shows the correlation between corrosion rates and material temperatures measured at a straw-fired stoker. Superheater temperatures in monofuel straw combustion, according to the chart, should be below 500°C in order to keep corrosion at a tolerable level. At superheater temperatures of 450°C or lower, significant

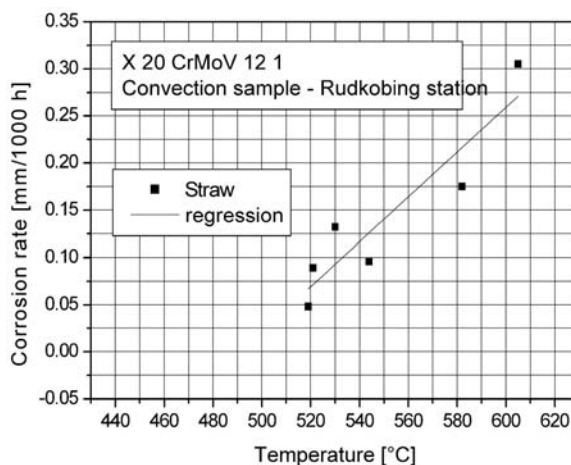


Fig. 6.10 Dependence of corrosion rate on material temperature (measured at a straw combustion plant by corrosion probe) (Clausen and Sorensen 1997)

corrosion does not occur in straw-fired stokers or cigar burner furnaces. If steam conditions above these temperatures are required, either a limited lifetime of the superheater has to be accepted or other fuels such as natural gas, wood or coal have to be used for superheating. A plant firing straw for the production of 470°C steam and firing wood for superheating to 540°C went into service in 1997 (Clausen and Sorensen 1997; Johnsen and Svendsen 1997). Severe corrosion can be avoided if austenitic steel (TP 347 FG, see Sect. 4.5) is used for the high-temperature section of the superheater. Inspections of corrosion at the Maribo Sakskobing plant (live steam temperature 540°C) indicated an expected lifetime of more than 25 years (Berg and Jensen 2008).

The important factors influencing the ash fusion behaviour are the concentrations of the alkaline earths Ca and Mg and the alkalis Na and K. Alkaline earths raise the fusion temperatures, while alkalis lower it. Chlorides, too, may cause the melting point to rise. Herbaceous biomass is the most likely to cause slagging on the grate, in the furnace and on the first superheater due to the high potassium content and the low ash softening temperature (Clausen and Sorensen 1997). The maximum furnace temperature should not be higher than 800–900°C (Obernberger 1996). Homogeneous fuel distribution, staged air injection to control the heat release, water-cooled grates, water- or steam-cooled walls and flue gas recirculation are means to limit the furnace temperatures and avoid slagging. In new straw-fired units precautions against slagging have been taken. A platen pendant superheater section is located above the furnace, operating as a slag condenser. The wet slag is designed to drip off the platen sections, and, because the spacing is so large, the slag deposits should not grow together to a massive slag formation (Berg and Jensen 2008).

In addition, fouling of the other heating surfaces is probable. Due to its smaller fly ash particles, herbaceous biomass poses a greater risk of fouling of the heating surfaces than woody biomass. Herbaceous biomass as a feedstock is finer and has a lower density and a higher alkali content than wood, which results in the smaller fly ash. In straw-fired plants in Denmark, the straw in some years had especially high chlorine and alkali contents, which caused various problems. Severe deposits were found throughout the entire furnace of both a grate and a cigar burner system, on radiant superheaters and air preheaters, in induced draught fans, ESPs and stacks. The problems could not be solved even by using soot blowers, so after 1–2 weeks of combustion operation with straw the plants had to be shut down and the heating surfaces cleaned (Obernberger 1996). However, satisfying boiler operating periods, inter-dispersed by manual cleaning intervals, are expected if pendant superheaters with large transverse pitches are employed.

In bubbling fluidised bed combustion, defluidisation of the bed resulting from the agglomeration of bed grains can be a major problem (Khan 2007; Khan et al. 2009). During biomass combustion in a fluidised bed, part of the ash and alkalis are released. Ash particles remaining in the bed can glue bed material grains together. This melt phase roughly matches the chemical composition of the ash. Another, even more severe type of agglomeration is the so-called coating-induced agglomeration (Visser et al. 2003). Sodium and potassium, released during combustion, form a very

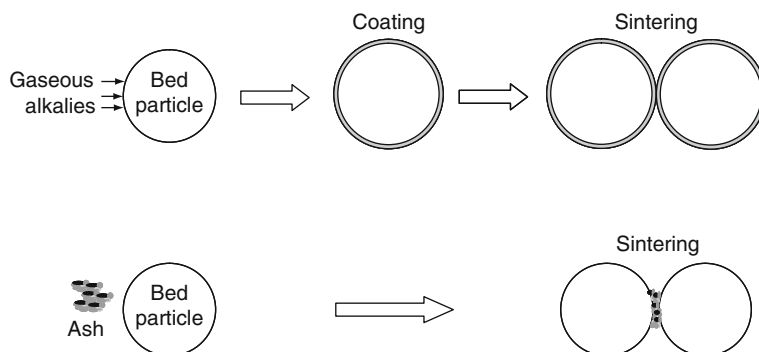


Fig. 6.11 Mechanisms of melt-induced and coating-induced agglomeration

thin and sticky coating of the bed grains. This type of agglomeration is believed to be the dominant process in commercial-scale installations. The mechanisms of melt-induced and coating-induced agglomeration are shown in Fig. 6.11.

High alkali fuels such as straw are known to provoke severe sintering of the bed, to an extent that the fluidisation can no longer be maintained, and operation must be halted. Fluidised bed technology is therefore limited in the fuels that it can fire alone – for instance, it is not suitable for the combustion of straw by itself. Investigations into straw combustion at a laboratory-scale furnace revealed that at a temperature of 800°C, defluidisation sets in after only 20 min. Tests with additives to prevent sintering were not very successful. The addition of coal was found to have a positive effect, though, suggesting a combined combustion of coal and straw in fluidised beds would be feasible (Lin et al. 1997; Bapat et al. 1997). In certain cases the use of quartz-free bed materials has also been successful in reducing sintering (Almark and Hiltunen 2005).

6.3 Biomass Gasification

The fundamentals of gasification are described in detail in Sect. 7.6 in the context of coal-based IGCC applications. Within this chapter, the focus is to discuss the special features of biomass gasification, though the principles of reactor design and gas cleaning are very much the same. The major differences in comparison to coal gasification are as follows:

- Due to the lower energy density and more limited availability of the feedstock, plant capacities only in the range of several megawatts up to 50–100 MW are considered. These sizes are much smaller than for coal gasification. The size of the plant is the dominating parameter for the power production costs and determines the choice of the gasification technology (the reactor design and the gasification medium). The preferred reactor technologies are fixed bed and

fluidised bed gasification, which, due to their lower gasification temperatures, give higher tar concentrations in comparison to the higher temperatures of entrained-flow gasification, the standard technology for coal gasification. For these capacities, gasifiers are usually air-blown.

- Biomass as a feedstock differs from coal. Biomass has a more inhomogeneous composition and particle size distribution, and generally fuel pre-treatment is required to produce a homogeneous feedstock suitable for the gasifier type. Hammer mills can only be used for relatively dry materials; to homogenise the size of wet materials, chippers equipped with blades can be used. Feeding must suit the higher bulk densities and the higher moisture contents that biomasses usually have, and therefore the feeding may require pre-drying. The higher volatile contents (in comparison to coal) are beneficial, requiring less residence time in the reactor for the conversion of the carbon, though the generally higher particle size may counteract this advantage. Alkali concentrations in the ash are higher than for coal.

6.3.1 Reactor Design Types

A large variety of gasification reactor designs are available for small and large scales. The designs can be classified in different ways (Kaltschmitt et al. 2009; Hofbauer 2007; Knoef and Ahrenfeldt 2005; Kaltschmitt and Bridgwater 1997; Spliethoff 2001; Kaltschmitt 2001; Higman and van der Burgt 2008; de Jong 2005), the most common being the following:

According to the gasification medium:

- Air-blown gasifiers
- Oxygen-blown gasifiers
- Steam gasifiers

According to the heat supply for the gasification:

- Autothermal or direct gasifiers: Heat is supplied by partial oxidation of the biomass, which results in a lower heating value of the product gas.
- Allothermal or indirect gasifiers: Heat is supplied from an external heat source, or for two-stage gasifiers, from the combustion zone to the gasification zone.

According to the pressure in the gasifier:

- Atmospheric
- Pressurised

According to the reactor design:

- Fixed bed
- Fluidised bed
- Entrained flow

The most common designs for biomass gasifiers are fixed bed and fluidised bed gasifiers. Fixed bed gasifiers are employed in the low capacity range of several MW_{th} ; fluidised bed installations are typically larger than 5 MW_{th} , though this technology is undergoing further development for use at capacities down to below 1 MW. Fluidised beds can be subdivided into bubbling and circulating systems. Technologies such as pressurised fluidised bed gasification, entrained-flow gasification and oxygen gasification are only economical at larger scales. Figure 6.12 gives the typical fuel capacity ranges that different gasifier designs are used at. The gas quality depends on the fuel quality, the gasification agent (i.e. steam or air) and the reactor design. The choice of the gasification agent has the dominating effect on the heating value of the product gas. Table 6.2 shows the various component fractions and the heating value of the product gas using either air (autothermal gasification) or steam (allothermal gasification) as the gasification agent. The reactor design has the dominating effect on tar concentrations (see Table 6.3).

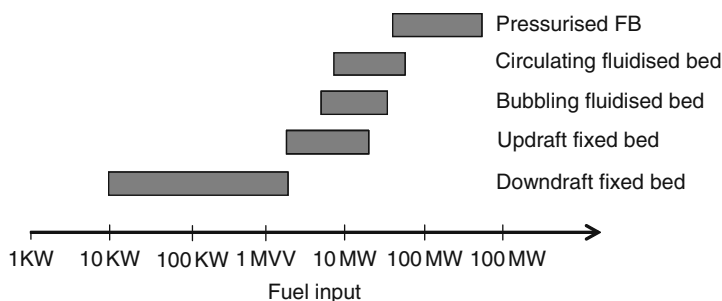


Fig. 6.12 Fuel capacity ranges for gasifier designs

Table 6.2 Heating value and product gas composition for air- and steam-blown gasification (Kaltschmitt 2001; FNR 2006; Knoef 2005)

Gas [vol.-%]	Gasification agent	
	Air	Steam
CO	10–20	25–47
H ₂	9–22	35–50
CH ₄	1–7	14–25
CO ₂	10–15	9–15
N ₂	40–55	2–3
LHV [MJ/Nm ³] ^a	3.5–6.5	12–17

^a Dry gas

Table 6.3 Tar and particle concentrations for different gasification systems (Kaltschmitt 2001)

	Fixed bed		Fluidised bed	
	Counter current	Co-current	BFB	CFB
	g/Nm ³			
Particle				
Range	0.1–3	0.2–8	1–100	8–100
Average	1	1	4	20
Tar				
Range	10–150	0.1–6	1–23	1–30
Average	50	0.5	12	8

6.3.1.1 Fixed Bed Gasifiers

In fixed bed gasifiers, gasification occurs in layers of the fuel bed, with different zones for the different gasification reactions that take place (pyrolysis, oxidation and reduction). A distinction (in respect to the flow) is made between counter-current and co-current gasifiers.

The most common type of counter-current gasifier is the vertical reactor, where the feedstock is fed from the top and the gasification agent added at the bottom. The directions of fuel flow and gas flow being opposed, separate reaction zones form in the reactor. The raw gas which is produced rises inside the reactor and leaves from the top, hence the common term “updraft gasification”. Counter-current gasifiers have the advantage of not requiring any special fuel preparation, thus allowing the gasification of a wide range of biomass types with different particle sizes and moisture contents.

Through forced convection, the gas heated by oxidation in the bottom zone rises and transfers heat to the fuel. The gas leaves the gasifier with a relatively low temperature, which reflects the high gasification efficiency of this process. The drawback is that the volatile matter, gasified in the pyrolysis zone, becomes part of the rising gas stream. In consequence, the raw gas of counter-current gasifiers contains a considerable amount of tar compounds.

In a co-current gasifier, the fuel and the gasifying agent move together in the same direction (Fig. 6.13). The pelletised bio-fuel first dries and pyrolyses in a near absence of air in the upper zones, then enters the very hot oxidation zone, where it is transformed into char and ash, and finally falls into the reduction zone. The gases, for the most part produced in the pyrolysis zone, are heated to a temperature appreciably over 1,000°C in the oxidation zone. In this process, the overwhelming majority of the high-tar gaseous compounds entering the oxidation zone are converted into low-tar components, which then react with the char in the subsequent reduction zone, producing additional gas. The raw gas issues from the bottom-most section of the reactor, hence the alternative term “downdraft gasification”. In contrast to counter-current gasification, the heat transfer between the bio-fuel and the gasifying agent in co-current gasification is small, so the raw gas has a relatively high temperature and the gasification efficiency is lower than that of updraft gasification.

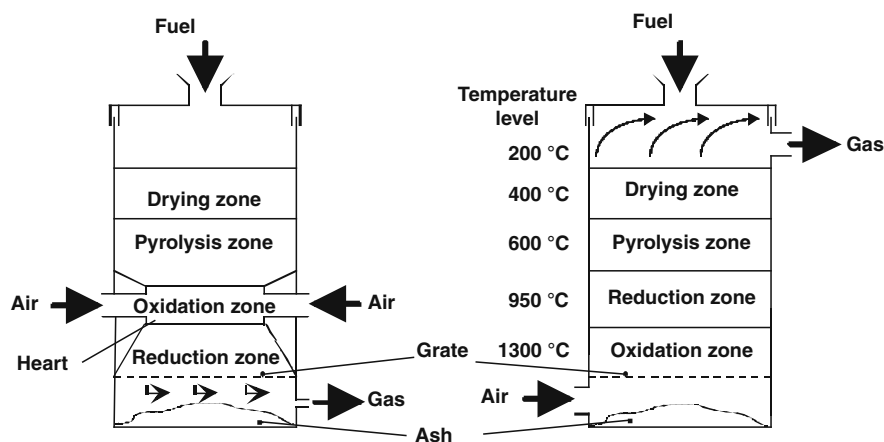


Fig. 6.13 Co-current gasifier (downdraft gasification, *left*) and counter-current gasifier (updraft gasification)

There is also a higher tendency for slag to form in co-current gasifiers than in counter-current gasifiers because of the high temperatures in the oxidation zone. A uniform temperature distribution within the individual reactor zones and a high permeability of the char to the gas are decisive factors for the gas quality. Co-current gasifiers therefore require a greater degree of fuel preparation to adjust the fuel particle size and the moisture content. The major advantage of co-current gasifiers is that the raw gas produced contains far less tar products than the gas from counter-current gasifiers.

Fixed bed gasifiers are generally atmospheric, air-blown (autothermal) and fuelled by wood, the latter of which results in a typical product gas heating value of between 4 and 6 MJ/m³. The gas can be used for heating purposes or in gas engines. Co-current bed gasifiers are superior to counter-current installations because of their lower tar concentrations, which are in the range of several 100 mg/m³. Quite a large number of fixed bed gasifiers have been tested, demonstrated or operated. However, most installations have had operational problems with fuel feeding, the gasifier itself, gas cleaning and prime movers. Current development concentrates on solving problems to do with automation, fuel feeding, the operation of the gasifier, gas cleaning and treatment of by-products. The tar concentrations are still a major concern.

A number of installations have been operated successfully in India for several years. These systems can be characterised by a near total absence of automation and by the use of wet gas cleaning with a sand bed filter for final tar removal. European standards, however, require fully automated installations and a gas cleaning system with disposable by-products.

Fluidised Bed Gasification

Fluidised bed gasification makes use of the advantageous mixing, reaction kinetics, gas – solid contact and heat transfer, as well as the ability to inject additives, of

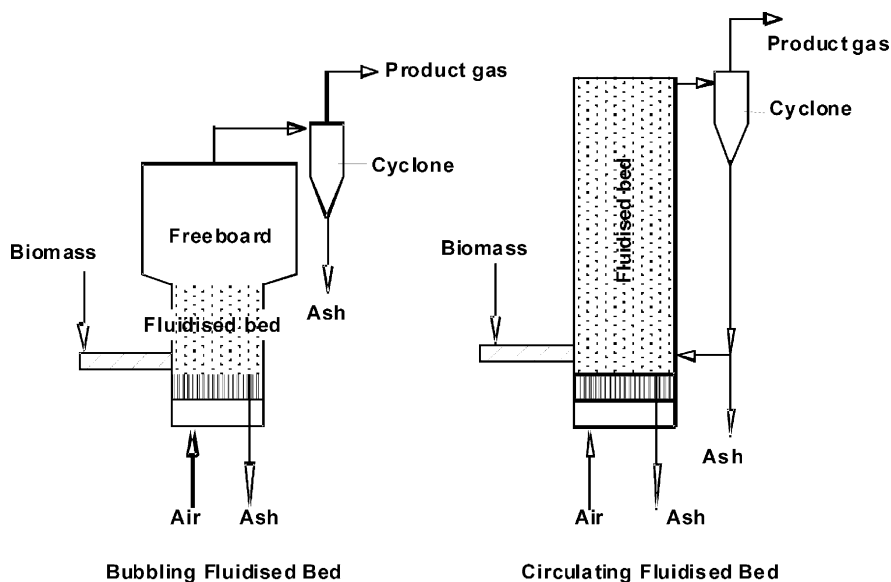


Fig. 6.14 Operating principles of fluidised bed gasifiers

fluidised beds. The bed material usually used is silica sand or, for high ash fuels, the ash of the fuel. The gasification temperature is typically around 800°C . The long residence time of the solid fuel and the intensive mixing are the reasons why very high gasification rates are achieved. The basic design types are bubbling (BFB) and circulating (CFB) fluidised beds, which can be seen in Fig. 6.14.

In a bubbling fluidised bed, the oxidant velocity is significantly lower than the terminal velocity of the bed material. The freeboard (the free space in the furnace above the bed) therefore has a gas flow which contains only small ash particles. In a circulating fluidised bed, in contrast, the oxidant approach velocity lies in the order of magnitude of the terminal velocity, and as a consequence the bed material is carried into the freeboard, thus forming a gas/solids flow through the entire reactor. By way of a cyclone, the transported bed material is separated from the gas stream and recirculated into the reactor. CFBs have significantly higher specific outputs. In addition, the gas/solids flow makes the mixing efficiency higher than in BFBs, which results in a better fuel conversion and lower tar contents. Drawbacks are the stricter fuel form requirements (the fuel needs to be in the form of grains) and the significantly higher pressure loss (meaning a higher process power consumption). What is more, controlling the bed material and the recirculation flow is more complex and the construction must be much taller than for a BFB. For lower outputs, a bubbling fluidised bed is therefore the better solution. With respect to the tar content, bubbling fluidised beds perform much worse (by about one order of magnitude) than co-current fixed bed gasifiers. Circulating fluidised beds are somewhat better, but do not reach the low concentrations of co-current fixed bed gasifiers.

In nearly all medium-to-large-scale electricity-producing biomass gasification demonstration plants to date, circulating fluidised beds (CFB) have been the preferred

technology. The main reasons are that CFBs can handle a high throughput, are easy to scale up and accept a wide range of fuels. However, tar conversion or tar scrubbing is required for all cold gas applications, that is, where the gas is fed to a boiler or an engine at ambient temperatures (Knoef and Ahrenfeldt 2005). Table 6.4 gives some examples of fluidised bed gasifiers in operation, but does not intend to provide a complete list. All CFB gasifiers are air-blown and produce a product gas with an LHV of 4–6 MJ/Nm³ (with wood as the fuel).

Pressurised gasification technology was successfully demonstrated from 1993 to 2000 in the world's first complete BIGCC (Biomass Integrated Gasification

Table 6.4 Medium-to-large-scale fluidised bed biomass gasification plants (Spliethoff 2001; Knoef 2005)

	Gasifier	Fuels	Use of product gas	Start of operation/ shutdown
Ruedersdorf, Germany	100 MW ACFB, Lurgi	wood, RDF, lignite, waste	Cement kiln	1996
Pietarsaari, Finland	35 MW ACFB, FW	Bark, Wood, waste	Lime kiln	1983
Norrundet, Sweden	27 MW ACFB, FW	Bark, wood waste	Lime kiln	1985
Rodao Mill, Portugal	17 MW ACFB, FW	Bark, wood waste	Lime kiln	1986
Zeltweg, Austria	10 MW ACFB, AEE	Wood	PC co-firing	1997–2000
Lahti, Finland	40–70 MW ACFB, FW		PC/NG boiler	1997
Geertruidenberg, Netherlands	80 MW ACFB, Lurgi	Wood waste	PC co-firing	2000
Ruien, Belgium	86 MW ACFB		PC co-firing	2002
Greve-in-Chianti, Italy	2 × 15 MW ACFB, TPS	RDF	Steam cycle 7 MW _e	1993
Burlington, USA	Battelle Columbus intercon- nected CFBs		Initially steam cycle power plant	1997/shut down
Hawaii, USA	IGT Renugas Pressurised BFB		Gas cleanup testing	1995–1997
Värnamo, Sweden (Bioflow)	18 MW PCFB, FW		IGCC	1993/1999
Tampere, Finland Biocycle	7 MW _e PBFB Carbona		IGCC	Unknown
Aire Valley, United Kingdom (ARBRE)	8 MW _e ACFB, TPS	Willow, poplar	IGCC	2001
Güssing, Austria	8 MW FICFB	Woodchips	Gas engine	2001

ACFB: atmospheric circulating FB, PBFB: pressurised bubbling FB, PCFB: pressurised circulating FB, FICFB: fast internal circulating FB, FW: Foster Wheeler, AEE: Austrian Energy & Environment, TPS: Termiska Processor AB

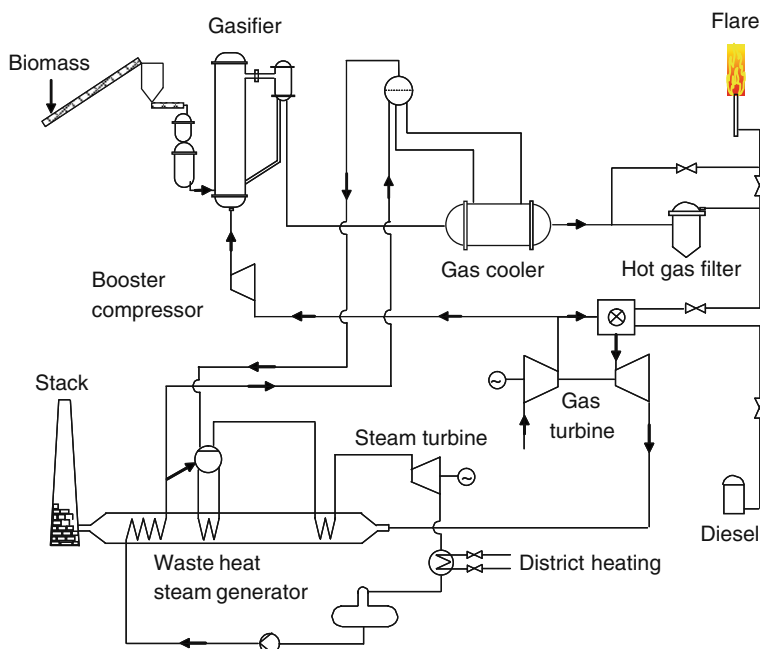


Fig. 6.15 Process flow diagram of the Värnamo plant (Kaltschmitt et al. 2009)

Combined Cycle) power plant in Värnamo, Sweden (Sydkraft 2001). The plant had a fuel input of 18 MW, an electrical output of 6 MW_{el} and a heat production of 9 MW_{th}. Figure 6.15 shows the process flow diagram of the Värnamo plant. Dried and crushed wood is pressurised in a lock hopper, then fed by screw feeders into the gasifier. The operating temperature of the fully refractory-lined air-blown CFB gasifier is 950–1,000°C and the pressure is about 18 bar. The gas produced in the gasifier is cooled to a temperature of about 350–400°C. After cooling, the gas enters a candle filter, where particles are removed. The clean product gas, with a heating value of 5–7 MJ/Nm³, is fed to the turbine. By the end of the demonstration programme the plant had been operated in gasification mode for more than 85,000 h and the gas turbine had run on product gas for more than 3,600 h. It was shown that pressurised BIGCC technology works. Experiences from the demonstration programme are given in Knoef and Ahrenfeldt (2005).

Two-Stage Allothermal Fluidised Bed Gasifiers

Allothermal gasification offers the advantage of producing a product gas with a higher calorific value than that from autothermal gasification. The principles of autothermal and allothermal gasification are described in Sect. 7.6.3. Allothermal gasification requires an external heat source to supply the required energy for the gasification process. Steam is used as the gasification agent. In order to avoid the

necessity of an external heat source for supplying the gasification heat, systems with two fluidised bed reactors have been developed. In one reactor, biomass is gasified, while biomass or char from the gasification reactor is combusted in the other. The transfer of heat from the combustion to the gasification reactor is accomplished by the transfer of sand, which acts as the energy carrier. Such a process has the advantage of producing a gas with a very low nitrogen content without the use of oxygen.

An example is the SilvaGas process, which has been demonstrated with a feed capacity of 44 MW in Burlington, Vermont, USA (see Fig. 6.16). The process consists of two CFBs. Biomass is fed to the gasification reactor, where it is mixed with hot sand and steam. Sand (the heat carrier) and the remaining char are separated in a cyclone and discharged to the combustor. The sand is reheated by burning the char with air in the combustor, then returned to the gasifier. The operating temperature of the gasifier is 815°C and the temperature of the combustor is 980°C (Knoef and Ahrenfeldt 2005).

The FICFB (fast internal circulating fluid bed) process developed by the University of Vienna is another example of such a process. It separates the steam gasification of the biomass from the combustion of the char, the latter of which is the heat source for the gasification. A commercial demonstration combined heat and power plant with a fuel power of 8 MW has been built in the town of Güssing, Austria. A schematic of the plant is given in Fig. 6.17. The synthesis gas is fired in a gas motor, generating 2 MW_{el} and 4.5 MW heat. The plant was put into operation in 2001. The gasifier operates as a bubbling fluidised bed with sand as the bed material. The sand and the ungasified char leave the bottom of the reactor and are transferred to the CFB combustor, where the char is burnt. The hot sand is separated from the flue gas in a cyclone and is returned to the gasifier, supplying the required gasification heat. The gasification reactor is operated at about 850–900°C, producing a gas with an LHV of 12 MJ/Nm³ (dry).

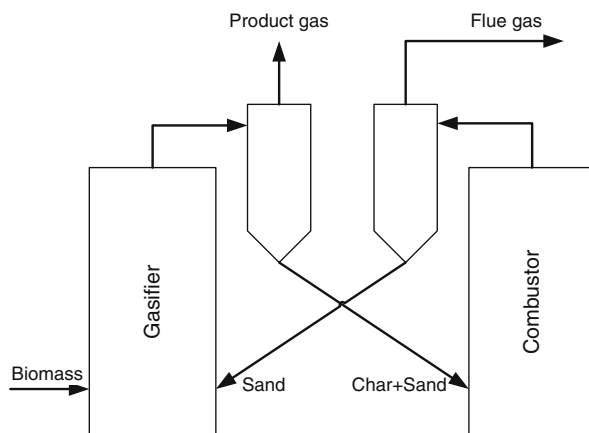


Fig. 6.16 Schematic of the SilvaGas (Batelle) gasifier

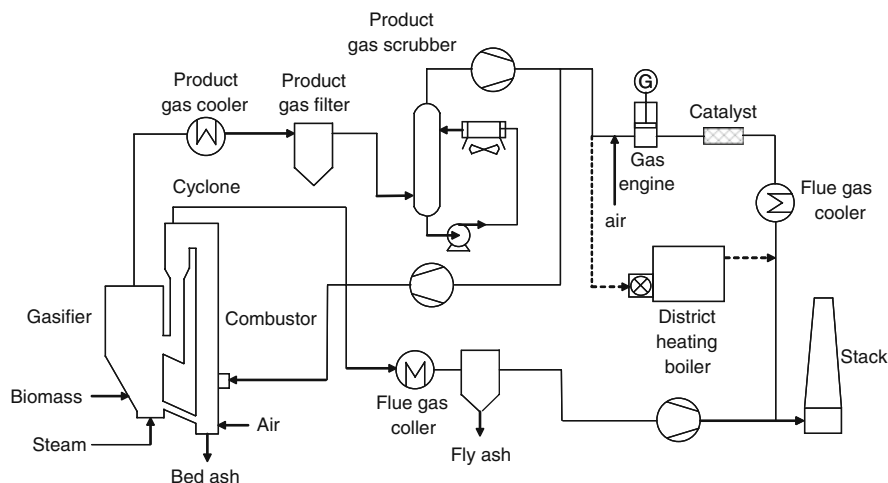


Fig. 6.17 Schematic of the Güssing plant (from Higman and van der Burgt 2008, © 2008, with permission of Elsevier)

Entrained-Flow Gasification

Entrained-flow gasification (described in more detail in Sect. 7.6) is the most common process for coal gasification. Despite the trend of using fluidised beds for biomass gasification, there are some cases of biomass gasification using entrained-flow technology.

One example is the Choren process, which is a combination of a low-temperature pyrolysis process with a two-stage entrained-flow gasifier, as shown in Fig. 6.18.

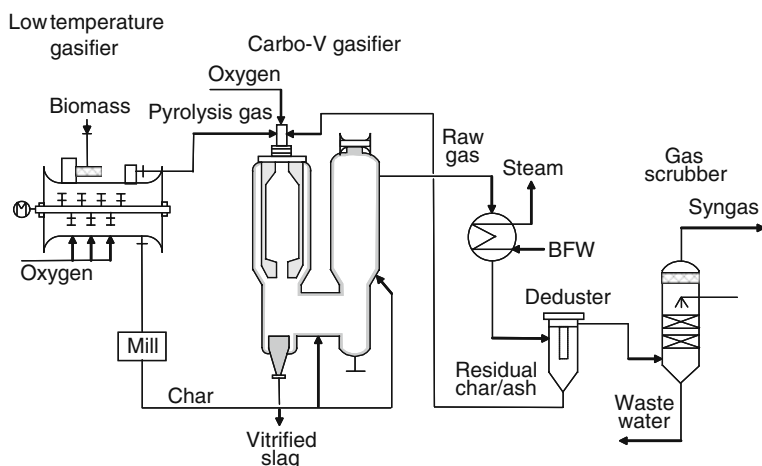


Fig. 6.18 Process flow diagram of the Choren process (from Higman and van der Burgt 2008, © 2008, with permission of Elsevier)

The biomass is fed into the stirred horizontal low-temperature gasifier and is pyrolysed in the presence of oxygen at a temperature of between 400 and 500°C. The pyrolysis gas and the char are extracted separately. The pyrolysis gas is then gasified in the first part of a high-temperature entrained-flow gasifier at temperatures of above 1,400°C, which effectively destroys the tars. Milled char from the pyrolyser is used as a chemical quench in the second part of the entrained-flow gasifier, where the hot gases from the first zone provide the energy for the gasification of the char. The raw gas leaves the reactor at a temperature of 800°C.

6.3.2 Gas Utilisation and Quality Requirements

The gas produced in the gasifier can be used in various ways for electricity production and for the production of process heat. These systems differ with respect to their efficiency, costs and gas quality requirements. The options for gas utilisation are shown in Fig. 6.19.

6.3.2.1 Gas Utilisation in Boilers and Cement Kilns

The most simple method of using gasification gas is to burn it in a steam generator or a cement kiln or to co-fire it in a coal power plant. This method requires only primary cleaning, such as by cyclones, of the gasification gas. If the gas is kept at temperatures above 500°C after it has been produced, tar conversion or tar scrubbing is not required. The majority of the commercially operated gasifiers supply gas for such thermal purposes. Among these, the plant at Rüdersdorf, Germany, with a thermal output of 100 MW, is the largest at present. The combustion of a gas instead of solid biomass simplifies the combustion process in the steam generator or the lime kiln (Rüdersdorf) and reduces ash-related restrictions. However there is no gain in efficiency over direct firing of biomass.

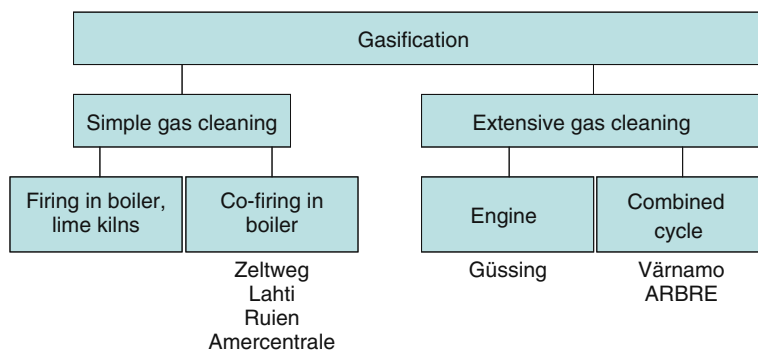


Fig. 6.19 Options for gas utilisation

Utilisation in Gas Engines and Gas Turbines

Gas utilisation in a turbine, motor or fuel cell for power production offers a higher efficiency than combustion in a steam generator.

The gas quality requirements for these applications are very high. In order to avoid fouling and deposits in the engine, the gas should be tar and dust free to a high degree. Typical concentration requirements for gas engines and gas turbines are listed in Table 6.5. Turbo-charged engines require an even higher gas quality for operation. It should be noted that the values given in published literature vary. Gasifiers available on the market today far exceed the indicated values when operated without gas cleaning. The removal of both tar and particles is therefore a prerequisite. The required concentrations are the result of a compromise between increased gas cleaning expenditure and a higher engine or turbine maintenance demand.

Engines are suitable for capacities between 50 kW_{el} and 10 MW_{el} and are used in connection with atmospheric fixed bed or fluidised bed gasifiers. With engines, the maximum electricity production efficiency is around 30%, although by including waste heat utilisation, the overall efficiency can be higher. Smaller plants yield lower efficiencies of up to about 25%. These efficiencies are somewhat above those that can be achieved by steam turbines in this capacity range.

From a capacity of about 5 MW_{el}, gas turbines are the better technology. The gasifiers suitable for use in connection with such turbines are atmospheric or pressurised fluidised bed reactors. With gas turbines, it is possible to increase the efficiency up to 45% by installing a tailing waste-heat boiler with a further steam turbine (capacities > 25 MW_{el}).

Only a few integrated gasification processes using gas turbines have been demonstrated, so experience with such plants is limited. In Vaernamo, Sweden, a pressurised fluidised bed furnace with an electrical output of 6 MW_{el} was in service from 1993 to 2000 (see Fig. 6.15). An atmospheric bubbling fluidised bed using the TPS (Termiska Processor AB) system, with an electric output of 8 MW_{el}, was put into service in 2000 (ARBRE Project, Great Britain). Both systems are now out of service due to economic reasons.

Table 6.5 Gas quality requirements for gas engines and gas turbines (FNR 2006; Spliethoff 2001; Kaltschmitt 2009)

	Gas engine	Gas turbine
LHV [MJ/Nm ³]	> 2	> 5
Tar [mg/Nm ³]	< 100	< 5
Particle [mg/Nm ³]	< 50	< 30
Particle size [μm]	< 3	< 5
Ammonia [mg/Nm ³]	< 30–55	
H ₂ S [mg/Nm ³]	< 1,150	< 1
Alkalies [mg/Nm ³]	< 50	< 3
Halogens [mg/Nm ³]	< 100	< 2

6.3.3 Gas Cleaning

Gas cleaning for biomass gasification remains an area of uncertainty, in need of development. Gas cleaning depends on the cycle arrangement and the gas quality requirements of the gas engine or gas turbine.

Atmospheric gasification systems require the removal of both particulates and tars. In atmospheric systems in which the product gas is fed to an engine or a compressor, cooling of the product gas down to about 50°C is required. In such cases, the most simple method for cleaning is to perform the separation of particles and tars in one step at the temperature that the gas is used at. Separation uses physical methods which require the cooling of the gas prior to cleaning, such as wet scrubbing or wet electrostatic precipitation.

The separate collection of tars and particles is a strategy applied in dry, high-temperature gas cleaning systems, using hot gas particle filtration and a thermal or catalytic cracker for tar reduction. Separate removal processes can also be advantageous when the tar separation unit, for example, a scrubber with an organic solvent at low temperatures, requires a low product gas particle content. Disposal restrictions, for instance, the forbidding of a mix of tars, particles and water to be discharged, may also necessitate separate removal processes.

Pressurised systems with gas utilisation in a turbine do not require a tar separation step if the gas flow can be kept above the dew point of the tars. In this case the chemical energy of the tars can be exploited in the turbine. Solids are removed in hot gas filters at temperatures of 400–500°C; typically, ceramic or metal candle filters with back-pulsing are employed. The temperature of the filtration step has to be below 500°C so that alkalis can be removed by condensation on the particles. Temperatures below 400°C have to be avoided to prevent tar condensation. Lower temperatures reduce the efficiency of the cycle because of losses of sensible heat and chemical energy in the tars (FNR 2006; Knoef and Ahrenfeldt 2005; Kaltschmitt 2001; Spliethoff 2001).

6.3.3.1 Tar Formation in Gasification

Tars are organic compounds (hydrocarbons), which condense at room temperature. The tar species formed in gasification are aromatic, heterocyclic aromatic or poly-aromatic hydrocarbons. The species that are commonly found in fractions above 5% are toluene, naphthalene and, when process temperatures are below 800°C, phenol. A great number of compounds occur only as trace elements, but taken as a group they can also constitute a considerable fraction of the tar quantity.

While the literature does not provide a uniform definition of the term “tar species”, efforts have been undertaken in European projects to standardise tar classification and tar measurements. Based on the so-called tar guideline (or “tar protocol”), a CEN (Comité Européen de Normalisation) standard (CEN Technical Specification) has been developed (Neeft et al. 2002; Good et al. 2005; Coda et al. 2004; DIN 2006). The intention is to make tar measurements in gases produced from biomass comparable.

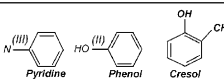
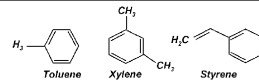
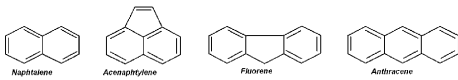
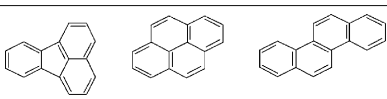

Class	Name	Example
1	GC undetectable tars. Heavy tars that condense at high temperature even at very low concentrations.	Gravimetric tars
2	Heterocyclic aromatic components. These components have a high water solubility.	 Pyridine Phenol Cresol
3	Aromatic (1-ring) components. Light hydrocarbons that are not important in condensation and water solubility issues.	 Toluene Xylene Styrene
4	Light polyaromatic hydrocarbons (2-3 rings PAH's). They condense at relatively high concentrations and intermediate temperatures.	 Naphthalene Acenaphthylene Fluorene Anthracene
5	Heavy polyaromatic hydrocarbons (4-5 rings PAH's). They condense at relatively high temperature already at low concentrations.	 Fluoranthene Pyrene Chrysene
Not considered as tar	Light hydrocarbons, that do not condense at ambient temperatures in normal concentrations	 Benzene

Fig. 6.20 Tar classification and chemical structure of selected tars. GC = gas chromatograph

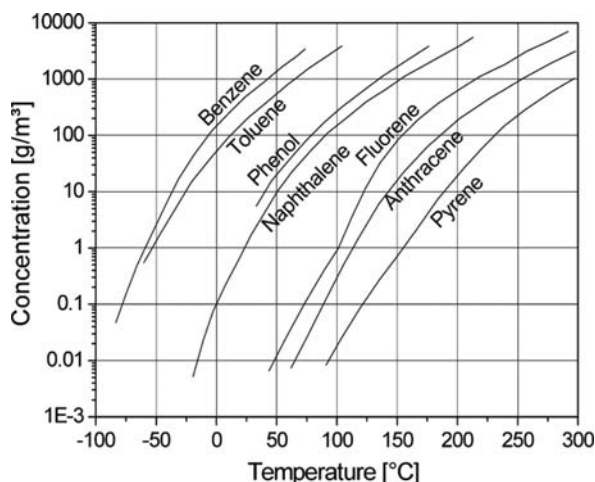
The “tar guideline” defines tars as the group of all organic compounds excluding gaseous hydrocarbons (C1–C6) and benzene. Benzene is not included because it does not condense at room temperatures at the concentrations typical of biomass gasification. The classification system in the “tar guideline” reports five different classes of individual tar compounds, as shown in Fig. 6.20 along with chemical structures of some typical tar components.

Class 1 consists of heavy poly-aromatic hydrocarbons (PAH) which cannot be detected with a gas chromatograph (GC) and is determined by subtracting the GC-detected tar fraction from the total gravimetric tar. Class 2 tars are aromatic compounds with hetero atoms (pyridine, phenol) while class 3 tars are light compounds with one aromatic ring (xylene, styrene, toluene). The last two categories, classes 4 and 5, consist of light polycyclic aromatic hydrocarbons (PAH) with two or three aromatic rings (naphthalene, fluorene, anthracene) and heavy PAH with four to five aromatic rings (pyrene) (Neeft et al. 2002).

Whether a component condenses when the temperature falls below the boiling point depends on the steam pressure and on the concentration. Figure 6.21 shows the saturation concentration of some typical tar components and benzene in nitrogen. Benzene at 25°C, for instance, reveals a saturation concentration of more than 300 g/m³. Since this value is higher than the typical concentrations in product gases by orders of magnitude, the condensation of benzene is not to be expected and benzene is not considered a tar. On the other hand, components such as fluorene have saturation concentrations of only a few milligrams per cubic metre at 25°C, so an almost complete condensation has to be taken into account.

The measurement of the dew point gives the highest temperature at which the first tar molecules can condense, information which is required to prevent a

Fig. 6.21 Saturation concentrations of some tar components in nitrogen (Spliethoff et al. 1998)



trouble-free operation of the gas cleaning train as well as trouble-free gas utilisation. The dew point is primarily dependent on the molecular mass of the compound and secondarily on the concentration of the compounds. An online tar dew point analyser has been developed based on the principle of tar condensation on an optical surface (van Paasen et al. 2005).

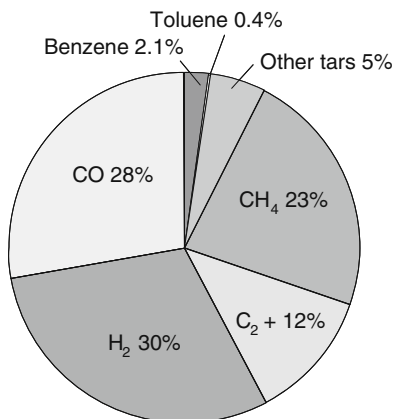
The tars grouped in each class of the “tar guideline” show a similar condensation behaviour. Class 3 tars have a very low dew point, much below ambient temperatures, so they will not condense in gasification systems. Class 5 tars have a high dew point (120°C at 0.1 mg/m³), so they will always condense.

Different methods, using different principles, exist for the measurement of the total tar concentration and the individual concentrations of tar compounds. The most widely used method is the tar guideline, which uses a modular sampling train in which the gas from biomass gasification flows through a series of impinger bottles filled with an organic solvent. Since the tars are diluted and collected in the bottles, they are analysed gravimetrically. Individual organic compounds can be determined by GC and high-performance liquid chromatography (HPLC). Another method traps tar vapours in a polypropylene cartridge using a solid amino-phase adsorbent. The analysis of the fractions is then made by means of a GC-flame ionisation detector (FID) technique (Braage et al. 1997).

A quasi-online tar measurement method has been developed based on continuous GC-FID measurements. Total hydrocarbons are measured both before and after removal of tars by condensation, the difference corresponding to the mass of the tars (Mörsch 2000). The application of laser spectroscopy for quantitative tar measurements is under investigation (Mitsakis et al. 2008).

The tar quantity and composition at the outlet of a gasifier highly depend on its construction and the operating parameters. Co-current fixed bed gasifiers produce a gas with a relatively low tar content, whereas the gas from counter-current gasifiers contains a high level of tar. In a co-current gasifier, the pyrolysis gases flow through

Fig. 6.22 Contribution of each gas component to the chemical energy of the product gas (beach wood, 800°C, $\lambda = 0.25$) (Mörsch 2000; Spliethoff et al. 1998)



the hot reaction zone, which results in the tar species getting cracked. The tar content of the product gas from fluidised bed gasification lies between these two values.

Investigations carried out in a bench-scale fluidised bed gasifier have revealed the impact of process conditions on tar formation. The reference case is the standard test with beech wood as the fuel, sand as the bed material, a gasification temperature of 800°C and an air ratio of $\lambda = 0.25$. The tar concentration in this case was 8.2 g/m³. Figure 6.22 shows the contribution of each of the gas components to the total heating value of the product gas of 5.2 MJ/m³.

Figure 6.23 gives the tar content of the product gas under variations of the operating and fuel parameters. The graphic shows for each case the tar concentrations after

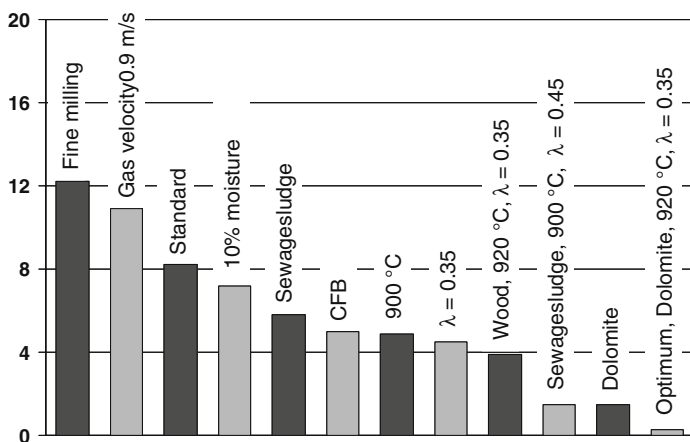


Fig. 6.23 Influence on the tar content of the tested operating parameters compared to the standard test case for a bench-scale fluidised bed (Mörsch 2000; Spliethoff et al. 1998)

a modification of a parameter in comparison to the reference case. With increasing process temperatures and air ratios, the tar content drops. At higher temperatures, there is also a shift in the tar composition towards lighter components. It should be noted that an electrically heated facility such as this bench-scale fluidised bed gasifier enables the air ratio and the temperature to be varied independently, whereas in a real autothermal gasifier, the air ratio and the gasification temperature are coupled. High temperatures in the bed can be limited by bed agglomeration.

The most effective measure to reduce tar concentrations established in the tests was the use of a catalytic material – in particular, dolomite. When employing this material, tar concentrations suitable for the direct utilisation of the gas are reached at temperatures above 900°C. The lowest tar concentrations using a dolomite bed, of just 250 mg/m³, were produced at 920°C and an air ratio of $\lambda = 0.35$. The resulting calorific value was 5.2 MJ/m³ at a cold gas efficiency of 85%. The main problem in using this technique is the low abrasion resistance of burned dolomite, meaning the catalyst gets entrained in the gas flow along with some of the bed material. To make up for this loss, it is necessary to constantly add a certain amount of bed material.

6.3.3.2 Secondary Tar Reduction

Tar cleaning is required if the product gas has to be cooled prior to use (for example, when used in gas engines) or if it has to be compressed (as in a combined cycle process with an atmospheric gasifier). The amount of tars present in the product gas, as described previously, depends on the gasification temperature, the reactor design and the biomass type. In most cases primary measures for tar reduction are not sufficient to meet the requirements for gas utilisation, so secondary measures for tar removal have to be implemented. These measures can take the form of a physical separation or a chemical conversion of the tars.

- In most cases physical methods are used for tar reduction. The two most common methods are scrubbing and electrostatic precipitation. Both methods remove the tars in a condensed form, which requires the gas to be cooled prior to the cleaning step. If the separated tars are not recycled to the gasification reactor, their chemical energy is lost from the process, reducing the cycle efficiency. Physical separation of tars and particles can be performed in a single step in one device at low temperatures.
- Chemical conversion of tars into lighter gas components can be achieved by thermal or catalytic tar cracking. After such processes, the chemical energy of the tars can be used to increase the heating value of the product gas.

Scrubbers

Wet scrubbing is the most common method for tar removal. If water is used as the scrubbing medium, the tar separation efficiency is limited and multistage cleaning

may be required. In order to be able to separate tars in a wet scrubber, the tars have to be condensed so that the aerosols and droplets collide with the water and increase their particle size. For this reason the product gas has to be cooled and saturated with water before the cleaning step. Because tars are hydrophobic and have a low solubility in water, tars which remain in the vapour phase cannot be removed.

Using a wet water scrubber to remove tar from the product gas requires a gas temperature of 30–60°C. Different washer types are in operation, for example, tower, rotating tower and Venturi washers. To achieve a higher tar removal efficiency, washers tend to be multistage, meaning a higher pressure drop. Clean gas concentrations of 20–40 mg/m³ can be achieved.

By using lipophilic liquids, which can act as solvents and so are used as a scrubbing medium, gas phase tars can also be removed. At the Güssing plant in Austria, a wet scrubber is used with oil (RME) as the scrubbing liquid. Used oil, saturated with tars and condensate, is vaporised and recycled to the gasifier. Another scrubbing technology (OLGA) has been developed and patented by ECN. The removal of tars is accomplished by scrubbing the tar-loaded product gas with a specially developed liquid oil in an absorption column. Tar removal efficiencies of 99% of the heterocyclic tars and almost all the heavy and light tars have been measured. The dew point of tars is as low as –17°C. The current design of the OLGA technology requires a dust-free gas.

Fixed Bed Filters

Fixed bed filters employing sand, saw dust or other materials as the filter medium are usually used in smaller units due to their simple design. Sand filters have a high removal efficiency of up to 95%, but use of them is problematic because of the need to dispose of the contaminated bed sand.

Wet Electrostatic Precipitators

A wet ESP, which is operated at temperatures of about 60°C, is a more attractive solution than a wet scrubber due to a higher removal efficiency, a lower pressure drop and a lower quantity of waste water being produced. The gas is cooled and saturated with water prior to the cleaning, then the liquid droplets and particles are separated by electrostatic precipitation. The removal of the condensed tars is improved by the use of a small water stream to flush them away. The ESP has the advantage of high particle and tar removal rates. The water – tar – dust mixture can be fed back into the gasifier.

Catalytic Tar Reduction

Higher tar reduction rates than those of physical gas cleaning can be achieved with catalytic hot gas cleaning at temperatures between 800 and 900°C. In this process,

the gasifier is backed by a “tar cracker” – a fixed bed, a fluidised bed or a honeycomb structure filled with catalytically active material. Materials that have revealed themselves to be extremely effective in reducing tar are limestone and dolomite. Nickel catalysts are also known for giving a very high reduction efficiency, being used in particular in commercial applications for steam reforming. The advantages of these systems are that theoretically, no waste matter is produced, and the chemical energy of the tar species remains in the gas, thus having an efficiency-enhancing effect. Catalytic hot gas cleaning is a feasible option for tar reduction if high-temperature fuel cells are used to exploit the product gas. In some investigations, dolomite has brought about tar reduction rates of more than 99.5%; nickel compounds achieved rates up to as high as 99.99%. The drawback of secondary tar crackers is the relatively high cost of the additional equipment. What is more, to raise the temperature to the optimum process level, a certain quantity of air often needs to be added to partially combust the gas, a factor unfavourably affecting the efficiency.

As well as using a catalyst in a secondary tar cracker behind the gasifier, it is possible to use one directly as part of the bed material in a fluidised bed gasifier. The choice of possible catalyst materials for this purpose is limited to limestone, dolomite and nickel compounds. Nickel catalysts, however, have a number of disadvantages if used as a bed material or additive. First, nickel dust is toxic – a particular concern for the ash, which gets contaminated by nickel. Second, the catalysts very quickly lose their effectiveness in the presence of carbon, as carbon deposits form on the surface of the nickel compounds. In consequence, nickel compounds are only suitable for application after a dust removal stage. The effectiveness of calcium-based sorbents for in situ tar cracking in a fluidised bed gasifier is shown in Fig. 6.23. In order to reduce the consumption of the catalytic bed material, the catalyst should have a minor attrition rate. A positive side effect of catalytic tar cracking is the reduction of ammonia to N_2 , with conversion rates of 70–80% (see Table 6.6).

Table 6.6 Removal efficiencies of different tar cleaning devices (Kaltschmitt 2001)

	Temperature [°C]	Particle reduction [%]	Tar reduction [%]	NH ₃ reduction [%]	H ₂ S reduction [%]	HCl reduction [%]
Sand filter	10–20	70–99	50–97	> 95	80–95	90
Washing tower/rotating Venturi washer	50–60	60–80	10–25			
Rotating sprayer	< 100	95–99	50–90		> 95	90
Wet ESP	40–50	> 99	0–60			
Bag filter	130	70–90	0–50			
Rotating filter	130	85–90	30–70			
Catalytic cracker	900		> 95	70–80		

Thermal Tar Reduction

Thermal tar reduction (cracking) offers an effective method for removal of tars and is typically performed at temperatures of about 1,200°C. The application of thermal tar crackers to fluidised bed gasifiers therefore would require a temperature increase from 800 to 1,200°C, which would have to be accomplished by partial oxidation. As partial oxidation would reduce the heating value of the product gas and would reduce the cycle efficiency, it is not used in biomass gasification.

Table 6.6 gives an overview of particle and tar removal efficiencies for different devices.

6.3.3.3 Particle Cleaning

The principles of particle cleaning are discussed in detail in Sect. 5.8 in the context of dust removal from combustion systems and in Sect. 7.4 in the context of pressurised fluidised bed combustion.

Low-temperature gas cleaning devices such as wet scrubbers or wet electrostatic precipitators have been described in Sect. 6.3.3.2. The systems operate at temperatures of about 50°C and are capable of removing both particles and tars. The particle removal efficiency is generally higher than that for tars.

Barrier filters are the common choice for dry particulate removal. Barrier filters with a rigid, porous, metal or ceramic candle can be operated at temperatures up to 500°C. However due to the higher costs and lower reliability, barrier filters such as bag house filters, which can be operated at temperatures of up to 350°C, are usually preferred. Depending on the dew point of the tars, condensation of tars in the filter can occur, blocking the filter. At the Güssing plant the fabric filter is operated at a temperature of about 160–180°C. Since some tar condensation on the filter does not inhibit the operation of the process as a whole, the filter is coated with a material that is cleaned from the filter together with the particulates when it is back-flushed. The cleaned material and particulates are fed back to the combustion section of the gasifier.

6.3.4 Power Production Processes

The focus of the following comparison of potential processes is power production at the medium to large scale. Within this range, circulating fluidised beds are considered the most promising technology, and therefore only these will be considered in this comparison, though the concepts can be transferred to other gasification systems. The processes under consideration are shown in Fig. 6.24.

- A. *Pressurised gasification with a combined gas–steam cycle*: The concept of pressurised fluidised bed gasification has been put into practice at the Värnamo Biomass Integrated Gasification Combined Cycle (BIGCC). Pressurisation

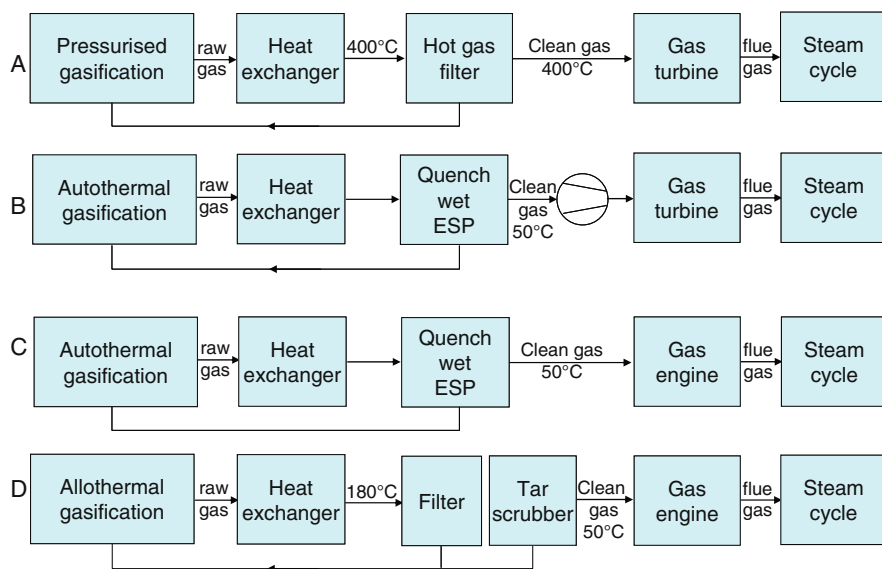
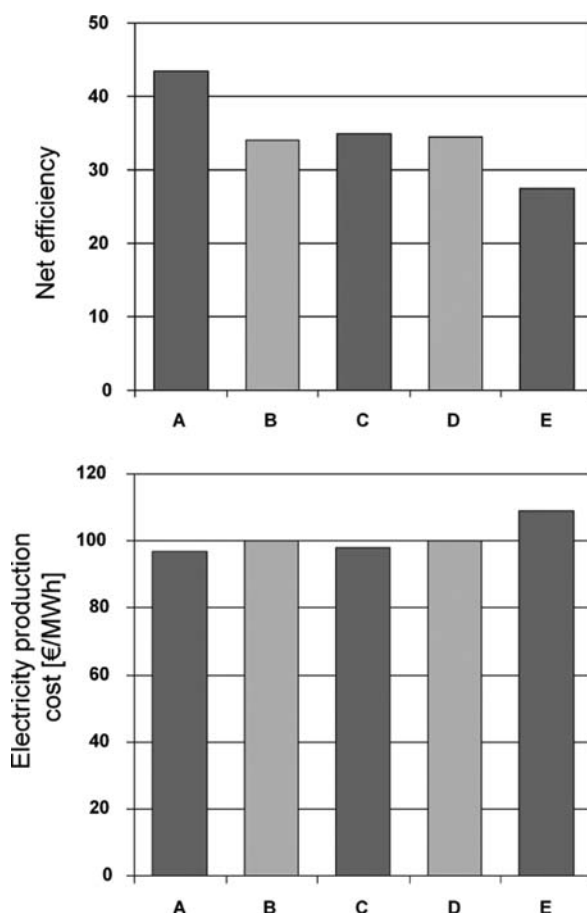


Fig. 6.24 Power production processes (Knoef and Ahrenfeldt 2005)

allows the use of the product gas in the gas turbine after gas cleaning without further compression. Therefore tar removal is not required and the heating value of the tars can be used in the gas turbine. The gas conditioning and cleaning consists of a heat exchanger to cool down the product gas to about 400°C and then dust removal in a hot gas filter. The cleaned gas is combusted in the gas turbine, which also supplies the gasification reactor with pressurised air. The waste heat of the gas turbine is transferred to a waste heat steam generator.

- B. Autothermal atmospheric gasification with a combined gas–steam cycle:* The product gas from the autothermal atmospheric gasifier is cooled in a heat exchanger down to a temperature of about 400°C. Additional cooling is achieved by quenching until a suitable temperature is reached for dust and tars to be removed in a wet electrostatic precipitator. The removal of tars is required for the compression of the product gas to the pressure of the gas turbine.
- C. Autothermal atmospheric gasification with a gas engine and waste heat utilisation:* The gasification and gas cleaning is the same as in case B, but instead of a gas turbine, a gas engine is used for power production. The flue gas from the gas engine has to be treated catalytically to achieve the CO emission limits. The waste heat is used in a steam cycle.
- D. Allothermal gasification with a gas engine:* This concept is based on the design of the Güssing plant. The allothermal fluidised bed gasifier produces a gas with a low nitrogen content, which is cooled, de-dusted in a bag house filter and cleaned from tars using a solvent scrubber.

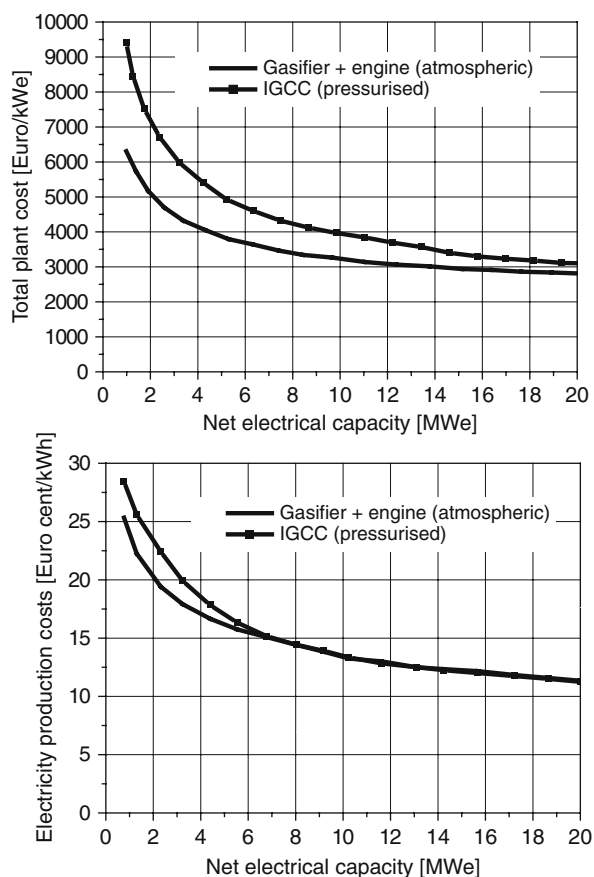
Fig. 6.25 Net electrical efficiency and production costs for biomass CFB processes (Knoef and Ahrenfeldt 2005)



The results of thermodynamic and economic assessments of the above process concepts are given in Fig. 6.25. The different gasification concepts are compared to a combustion-based conventional steam cycle (case E). The assessment has been carried out for an electrical output of 20 MW_{el}; details can be found in Knoef and Ahrenfeldt (2005). Pressurised fluidised bed gasification yields the highest efficiency of 44%, whereas the other cases achieve efficiencies of about 35%. The lower efficiency of atmospheric gasification with a gas turbine is due to the power demand for the compression stage. All gasification processes achieve much higher efficiencies than the 28% of the reference steam cycle.

The importance of the scale is demonstrated in Fig. 6.26, which shows that the pressurised process has lower power production costs only above a certain scale (6 MW_{el}), because of the higher capital costs of the pressurised system.

Fig. 6.26 Capital and electricity production costs as a function of the capacity for biomass CFB processes (Knoef and Ahrenfeldt 2005)



6.4 Thermal Utilisation of Waste (Energy from Waste)

Waste is an unwanted or undesired material or substance. The European Union, under the Waste Framework Directive (EU 2008), more precisely defines waste as an object the holder discards, intends to discard or is required to discard. Thermal waste treatment is an important element in modern waste management.

Table 6.7 lists the amount of various wastes fired in thermal waste treatment plants in Germany in 2006. The rise of thermal waste treatment plants was largely driven by the regulation restricting the disposal of MSW (AbfAbIV 2000; TASI 1993). Today thermal waste treatment fulfils several purposes:

- To destroy, convert, separate, concentrate or immobilise harmful or hazardous components of the wastes
- To reduce the volume and amount of waste as far as possible

Table 6.7 Thermal treatment of waste in Germany in 2006 (Statistisches Bundesamt 2008)

	Amount treated in 2006 (millions of tonnes)
Municipal solid waste (MSW)	18,142
Sewage sludge	1,669
Hazardous waste	1,056
Waste used as a fuel in power/CHP plants and others (wood residues, RDF, paper)	12,330

- To transform the remaining residues into usable substances or to put them into a disposable form
- To use the thermal energy released in the process to the greatest extent possible

Until 2008, the distinction between thermal treatment being “disposal” or “recovery” was made in Germany using the heating value of the waste, according to the German waste law “Kreislaufwirtschaftsgesetz” (KrWG 1994). The incineration of waste with heating values above 11 MJ/kg was considered “recovery”, whereas below this value, the process was considered “disposal”. The distinction followed a similar pattern in Italy, where the utilisation of so-called CDR (combustibles from waste, “Combustibile derivato dai rifiuti”), with heating values of more than 14 MJ/kg, was considered recovery. This kind of division seems to be rather arbitrary or politically motivated and is not consistent from the engineering point of view. The efficiency of an energy from waste (EfW) system depends on numerous factors, such as the process technology, the energy consumption of pre-treatment and transportation and generally speaking whether heat or electricity is used in the process. The heating value in itself only plays a minor role.

The European Waste Framework Directive (EU 2008) of 2008 introduced a new methodology for distinguishing between “disposal” and “recovery” with respect to thermal treatment:

- “Recovery” means “any operation the principal result of which is waste, serving a useful purpose by replacing other materials which would otherwise have been used to fulfil a particular function, or waste being prepared to fulfil that function, in the plant or in the wider economy.” In plainer English, recovery is a process in which waste replaces another material or where it is prepared so that it can replace another material. The recovery criterion (R1-recovery) based on energy efficiency, described in Annex II of the directive, is explained below.
- “Disposal” means any operation which is not recovery, even where the operation has as a secondary consequence the reclamation of substances or energy. The disposal criterion (D10-treatment) is described in Annex I of the directive.

In practice, the distinction between “disposal” and “recovery” is made depending on the energy efficiency of the process. Plants gaining approval after 2008 need to

have an energy efficiency corresponding to an *R1* criterion above 0.65. The *R1* criterion is defined in the footnote of Annex II of the Waste Framework Directive (EU 2008) as follows:

$$R1 \text{ energy efficiency} = (E_p - (E_f + E_i)) / (0.97 \times (E_w + E_f))$$

In which

- E_p is the annual energy produced as heat or electricity. It is calculated by multiplying the electrical energy by 2.6 and the heat produced for commercial use by 1.1 (GJ/year)
- E_f is the annual energy input into the system from additional (i.e. fossil) fuels contributing to the production of steam (GJ/year)
- E_w is the annual energy contained in the treated waste calculated using the net calorific value of the waste (GJ/year)
- E_i is the annual energy imported into the process excluding E_w and E_f (GJ/year) (i.e. the auxiliary power consumption)

In a simplified way an *R1* of 0.65 corresponds to a gross power generation efficiency of 24%, if only electricity is produced and no fossil fuel is used. This also assumes a perfect availability of the power generation components over the year (i.e. no unscheduled shutdowns and operation at full capacity) as well as an electricity in-plant consumption of 3% (referring to the gross heat input).

EfW plants in Switzerland and Amsterdam do not use fossil fuels, as assumed above. In most other plants in Europe, Japan or the USA, fossil fuels are used for start-up, shutdown and during unstable combustion conditions (for instance, as soon as the furnace temperature drops below 850°C at the 2 s-level of control). The use of, for example, 2.5% fossil fuel (by heat input) would decrease the *R1* by 0.04. For this reason, it is estimated that in some cases an efficiency at the design point of up to 28% is necessary for reaching, over the year, an *R1* value of 0.65 in practice.

The topic of energy efficiency in EfW systems is further detailed in Sect. 6.4.6.

One of the particularities of using waste as a fuel in combustion systems for energy generation is the need to meet the special emission requirements for both the flue gas and ash. In Europe, these emissions are subject to strict limits laid down in the Waste Incineration Directive (EU 2000), which shall be replaced in 2009 by the Industrial Emissions Directive (IED). The limits of the Waste Incineration Directive can be found in Table 5.7.

There are numerous kinds of waste materials, each of which needs to be treated in an environmentally sustainable manner. The focus in this chapter will be on municipal solid waste (MSW), refuse-derived fuel (RDF) and sewage sludge. More information on these waste streams can be found in Sects. 2.2.1.2, 2.2.1.3 and 2.2.1.4. They are produced in significant quantities and have properties such that it is reasonable to use them for energy recovery. Because half of the energy contained in MSW is biogenic, EfW is considered an important additional renewable energy source.

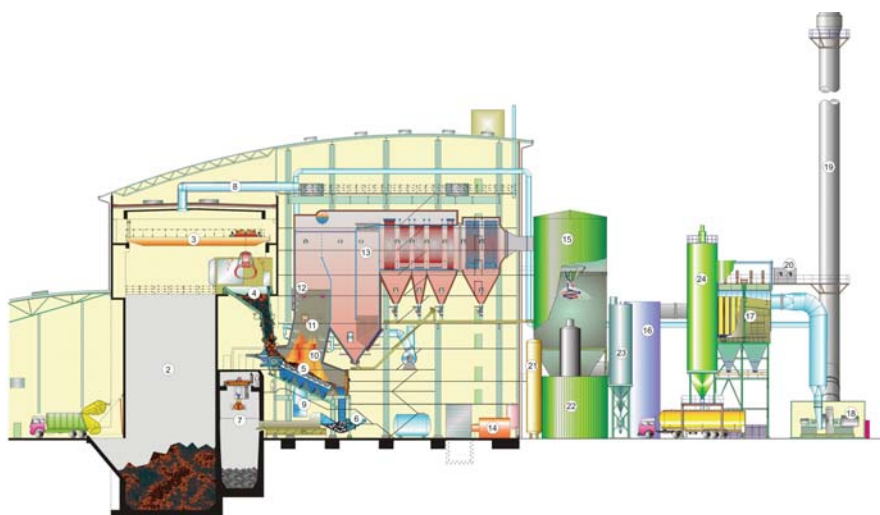


Fig. 6.27 Classical EfW system suitable for MSW, RDF and the co-combustion of sewage sludge
(Source: Martin)

Municipal Solid Waste (MSW): In Germany, the thermal treatment of MSW and the resulting energy generation in classical EfW plants uses the most waste, with about 18 million tonnes being treated in 70 plants. An example of a typical plant employing classical technology is shown in Fig. 6.27. The key components are a waste bunker, a waste feeding process, a grate-based combustion system, a boiler, a turbine and a flue gas cleaning process. Typical boiler steam conditions are 380–420°C and 40 bar. The flue gas cleaning system is based on the injection of lime and activated carbon to remove pollutants and a downstream baghouse filter for removing the fly ash and the products of reaction from the pollutant removal. NO_x is reduced by ammonia injection in the furnace.

This type of EfW system is typically used for untreated municipal solid waste (also called residual waste), which can be quite heterogeneous in composition. In addition, residues from mechanical – biological waste treatment plants, as well as industrial wastes, are also fed to classical EfW plants. On top of that, co-combustion of between 5 and 15% sewage sludge (by mass input) is quite common.

Refuse-Derived Fuel (RDF): RDF (also called solid recovered fuel (SRF) or substitute fuel) is produced from municipal solid waste by mechanical – biological treatment, in most cases including commercial waste as a feedstock. RDF usually has a higher heating value and a narrower particle range than untreated MSW, which makes it possible to use it in fluidised bed combustion systems, unlike untreated MSW. Another option is co-combustion in coal power plants or in cement plants. For RDF processed from MSW it is most common to use grate-based combustion systems.

Sewage Sludge: Sewage sludge is a type of waste which arises in a relatively homogeneous form, largely because of its high water content. Typical sewage

sludge, even after digestion and mechanical drying, has a water content of between 60 and 80%. It can be co-combusted in this state in waste incineration plants and brown coal-fired power stations or used as the only fuel in fluidised bed systems. For co-combustion in hard coal-fired power stations, thermal drying is usually necessary. In 2004, more than one-third of the 2.2 million tonnes of dry sewage sludge matter produced was combusted in Germany: 0.4 million tonnes in 21 monofuel combustion plants and 0.3 million tonnes in 25 coal power stations, with 0.05 million tonnes being co-combusted in EfW plants (Quicker et al. 2005; Schmelz 2006; Hermann 2004).

The following subchapters cover

- the historical development of EfW systems (with a focus on MSW, being the most used waste source for energy generation),
- grate-based combustion systems (most common for MSW and also used for RDF and the co-combustion of sewage sludge),
- pyrolysis and gasification systems (used for MSW and RDF),
- RDF systems (combustion systems being designed specifically for RDF using suspension or fluidised bed combustion),
- sewage sludge systems (mono-combustion),
- boilers for MSW, RDF and sewage sludge as a group and
- flue gas treatment for MSW, RDF and sewage sludge as a group.

The co-combustion of RDF and sewage sludge in coal power plants is described in Sect. 6.5.

6.4.1 Historical Development of Energy from Waste Systems (EfW)

The incineration of MSW began around the end of the 19th and the beginning of 20th centuries, when the first plants were constructed in England, the USA and Germany. The main purpose of these plants was to promote public health, as the practice of distributing waste on fields had been instrumental in the spread of cholera epidemics. The technology was very simple: brick-lined cell ovens with a fixed metal grate over an ash pit below, with one opening in the top or side of the oven for loading and another opening in the side for removing the solid residues, clinker or ash. Since that time, the technology has made huge advances in terms of emission control and energy efficiency (Gohlke and Spliethoff 2007).

An efficient energy from waste industry developed in Europe in the 1960/1970s and in the USA in the 1980/1990s. Typical EfW plants constructed in the USA during this period used steam parameters of 60 bar/443°C and, generally speaking, were more energy efficient than the European 40 bar boilers or the Japanese 20 bar boilers of this time. Waste incineration or energy from waste plants is often described in the USA as waste-to-energy plants (WTE), which can be considered as essentially the same.

In the 1980s, waste incineration plants became *the* symbol of environmental contamination: citizens were opposed to the throw-away society and “dioxin spouting” on the outskirts of cities. One of the reasons for the protests against waste incineration was the discovery of polychlorinated dibenzo-dioxins and polychlorinated dibenzo furans (PCDD/F), often simply called “dioxins”, in the flue gas of waste incineration plants. Even though these dioxin concentrations in the flue gas were comparatively low, a connection was made to the dioxin accident at the chemical plant in Seveso, Italy, in 1976.

The protest led, in Germany, for example, to important developments: by 2005 more than half of all household waste (55%) was recycled as bio-waste, waste paper, waste glass or packaging waste. Since June 1, 2005, untreated waste has no longer been sent to landfill. Furthermore, because of stringent emission regulations, waste incineration or EfW plants are no longer significant in terms of emissions of dioxins, dust or heavy metals. This is the case even though total waste incineration capacity has almost doubled since 1985 (UBA 2005b). Table 6.8 shows the historical development of waste treatment capacity in classical EfW plants in Germany.

Regardless of the European or international increase in total EfW capacity, cost reduction remains one of the main targets for the development of EfW systems in Europe, because it is necessary to compete with the very cheap alternative of landfill. The inexpensiveness of landfill will obviously only remain as long as the long-term environmental burden is not considered. In some of the more advanced European countries, landfill of MSW is restricted and efforts are being concentrated on further improving the energy efficiency of EfW plants (Bonomo 1998; Van der Linde 2003; Wandschneider 2005; Fischer 2005; Seguin 2004). The driving force behind the implementation of high-efficiency systems is usually a premium for renewable electricity, for example, up to 170 €/MWh under the CHP6 program in Italy (which has now ended) (Pfeiffer 2003). In countries with sophisticated waste management systems like Sweden, Denmark, the Netherlands, Belgium, Germany and Switzerland, more than 170 kg of MSW per person per year is used for energy generation. Consequently, less than 40 kg per person per year is sent to landfill in these countries.

Table 6.8 Historical development of total waste treatment capacity in classical EfW plants in Germany (UBA 2005b)

Year	Number	Capacity, in 1,000 tonnes per year
1965	7	0.718
1970	24	2.829
1975	33	4.582
1980	42	6.343
1985	46	7.877
1990	48	9.2
1995	52	10.87
2000	61	13.999
2005	66	16.9
2007	72	17.8

The development in the USA was different: in 1994, the US Supreme Court rejected the practice of waste flow control (US Supreme Court 1994). Waste flow controls are legal provisions that allow state and local governments to designate the places where municipal solid waste is treated or disposed. This rejection then led to an abrupt halting of the development of energy from waste schemes. Instead, waste was transported by truck to mega-landfills which had been developed at low cost in more remote areas. This resulted in the very negative environmental impacts of landfilling: land consumption, transportation by truck over long distances, methane emissions, leaching of pollutants to soil and to groundwater. The development of more sustainable waste management systems was no longer possible. However, after landfill taxes and premiums for renewable electricity were introduced and fuel prices for transport by truck increased, there has been an increasing trend towards the development of improved EfW processes since the middle of the first decade of the 2000s.

Waste incineration is very widespread in Japan, as landfill has traditionally not been an option due to the lack of space. In the 20th century most waste was incinerated in numerous small plants. In the year 2000, there were still more than 10,000 plants with a capacity of less than 2 t/h, while there were more than 2000 plants with a capacity of more than 2 t/h (Konda 2000). The UNEP (United Nations Environment Programme) published a study in 1995 which showed that 4 kg of the global 10 kg TEQ (toxic equivalent units, a measure of all toxic dioxins, furans and PCBs in terms of the most toxic form of dioxin, 2,3,7,8-TCDD) of dioxin emissions per year came from Japan. This resulted in a major policy change and a new dioxin regulation was put into effect in 2003. The dioxin emissions for existing plants larger than 4 t/h were limited to 1 ng TEQ/Nm³ and for new plants to 0.1 ng TEQ/Nm³. Up until the end of 2002, a dioxin emission of 80 ng TEQ/Nm³ was still acceptable. As a consequence, most small incineration plants had to shut down, because retrofitting dioxin reduction technologies was not economically feasible. The formation and destruction of dioxins is discussed in Sect. 6.4.8.

These new regulations and the general policy from 1995 to 2005 in Japan were drivers for the development and installation of new gasification and ash melting processes. Today, there are more than 90 gasification and over 90 ash melting processes in operation or under construction in Japan. The main purpose of these processes is to improve ash quality (by reducing the extent of leaching of the contaminants from the ash) and to reduce the total dioxin output per tonne of waste. This development peaked in 2000, when a total capacity of 3 million tonnes per year of plants were ordered. Half of this capacity was realised through gasification plants and the other half through stoker-based systems combined with ash melting (Vehlow 2006). However, both these processes are considered to be somewhat inadequate in terms of energy/cost efficiency and availability (Martin et al. 2005). The same experience was had in Europe in the 1990s with gasification systems, which have now all been shut down. For this reason, a policy change was announced in Japan in 2005. The new policy states that gasification or ash melting would no longer be required if plants were situated in remote areas or if it could be proved that the plants had at least 15 years of landfill capacity for ash or other ash recycling options. As a consequence,

development has once again focused on grate-based processes with an improved ash quality (for example, by oxygen enrichment, bottom ash washing, reduction of dioxin from the fly ash and the use of bottom ash in the cement industry).

6.4.2 Grate-Based Combustion Systems

The classical EfW system, a grate-based system, is shown in Fig. 6.27. They are widely used for MSW, RDF and the co-combustion of sewage sludge. In the USA, this kind of grate-based combustion systems for MSW are also called “mass-burn” combustion. Fluidised bed or suspension firing plants, which are specially designed for RDF, are described in Sect. 6.4.4. Section 6.4.5 deals with systems for the mono-combustion of sewage sludge.

Grate-based EfW systems are offered by a great number of manufacturers, though the principal processes differ only slightly from one another. Such a plant consists of the following functional sections:

- Waste unloading, storage and pre-treatment
- Feeding and combustion
- Waste heat recovery
- Bottom ash (also called slag) removal and residual material treatment
- Flue gas cleaning (Bilitewski et al. 2000)

Figure 6.28 shows in more detail the schematic of the combustion system, which is the core component of EfW systems. The delivered waste is stored in the refuse pit at first. There, in order to ensure an even fuel quality, it gets mixed and homogenised by means of cranes. Via a feed chute and special feeding and charging devices, the

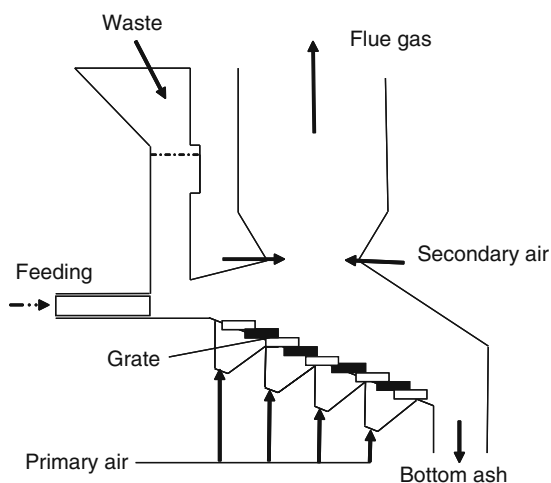


Fig. 6.28 Schematic drawing of a grate-based combustion system for MSW

waste is loaded onto the stoker. The grate transports the fuel through the various zones of the combustion process while providing a good mixing. On the whole, the waste remains on the grate for about an hour until all processes are completely finished. The remaining ash is discharged into a basin filled with water, called the ash discharger. Subsequently, the cooled bottom ash can be subjected to various treatments, such as the separation of metals, crushing and sieving, to make it more suitable for use as a filler material.

The hot flue gases are cooled in a steam boiler, creating steam, which is normally used for electricity production. Additionally, it can be used to deliver heat for district heating. In EfW plants, the design of the steam generator has to be given special attention, because the flue gases from the combustion of waste are very aggressive, reducing the heat transfer through fouling and slagging and potentially corroding the heat transfer surfaces. For this reason, the allowable steam conditions in EfW plants are markedly lower than the conditions in large steam power plants – a fact contributing to the lower power generation efficiency of a waste treatment plant. The design, the restrictions on it and the possible means to increase the efficiency are discussed in Sect. 6.4.7.

The cooled flue gases are treated in a flue gas cleaning section, where the remaining fly ash, including components such as heavy metals and organic toxic components, e.g. dioxins, is removed. Additionally, gaseous pollutants such as hydrochloric acid, hydrofluoric acid, sulphur dioxide and nitrogen oxides are removed. A large variety of technologies and systems are available for this purpose and are discussed in Sect. 6.4.9.

For the combustion of untreated municipal solid wastes, the grate furnace is the most widespread technology employed in the large EfW markets of Europe, the USA and Southeast Asia. The essential advantage of the grate furnace is its capacity to process lump-sized waste and accommodate fluctuating fuel properties (Bilitewski et al. 2000; Scholz et al. 2001; Doležal 1990; UBA 2005a; Effenberger 2000; VDI 2006, 2007; EIPPCB 2005). In a grate furnace, the combustion process can be divided into different phases, though these overlap to a large degree:

- *Drying and devolatilisation:* In the upper area of the grate, the waste is heated, dried and devolatilised by heat radiation from the refractory-lined walls or ceilings or by thermal convection of hot flue gases. The devolatilisation process usually takes place in a reducing atmosphere.
- *Gasification and combustion:* In these processes, two kinds of reactions run concurrently – heterogeneous fuel reactions and homogeneous reactions of the released volatile components in the combustion zone above the grate.
- *Burnout zone:* In the final area of the grate, complete burnout should be achieved. The burnout of bottom ash in modern waste incinerators is characterised by a loss of unburnt carbon of less than 2%.
- *Secondary combustion:* For the complete burnout of CO and fly ash particles, there is always a secondary combustion zone. Here, air or recirculated flue gas is added to complete the combustion.

The primary air injected below the grate provides the combustion air for the combustion reactions on the grate; the remaining combustion air is added as secondary air above the grate for the post-combustion of the unburnt components. In order to be able to adapt the primary air injection to the requirements of the individual combustion zones, the primary air ducts are divided into compartments. Large grate furnaces can have compartments not only in terms of length but also in terms of width. The amount of air per compartment can be controlled. With the aim of a complete burnout at low NO_x emissions in mind, the combustion of the bulk is mostly set at near-stoichiometric to air-deficient conditions. To give an example, at a total air ratio of about 1.6, 60% of the air is, as primary air, divided up and distributed to the drying zone (10%), to the combustion zone (40%) and to the burnout zone (10%). The rest of the total air is secondary air. The grate bars are protected and cooled by the primary air and the ash layer. The combustion air can be preheated for low waste LHV's. For high LHV's, e.g. above 12 MJ/kg, no preheating is applied and for LHV's of 6 MJ/kg, the preheating temperature can be up to 150°C. Forward pushing grate systems may need water cooling in case of high LHV's.

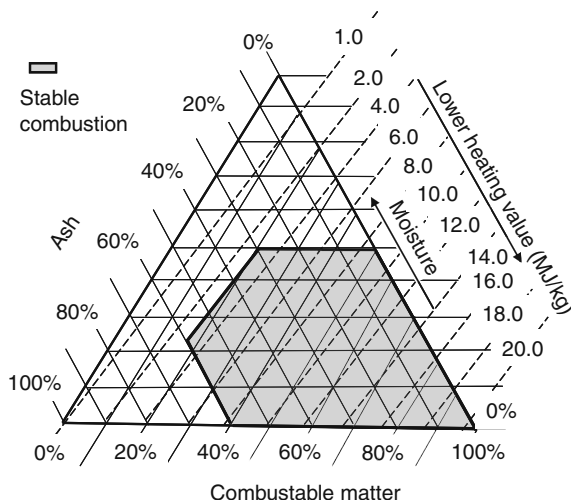
At the end of the grate, the bottom ash falls into an ash discharger, which is effectively a water basin, and is cooled down for further treatment. The amount of bottom ash is approximately 25% by weight or 10% of the volume of the original waste. A small portion of the fine ash may fall through the grate. These riddlings (also called grate siftings) are collected and transported pneumatically or via mechanical conveyers to the ash discharger.

The maximum combustion temperatures along the grate should be set so that the mineral components on the grate become sintered while avoiding the formation of large agglomerates, the latter because they would hamper the passage of air and thus the combustion process. Such an incidence is counteracted regardless, however, by the stoking effect of the grate firing system. Generally speaking, waste bed temperatures between 800 and 1,250°C are desired. The waste bed temperature can be monitored by infrared cameras in the furnace roof. A few plants have been constructed with oxygen enrichment of the primary air in order to achieve an improvement in the ash quality through the resulting higher waste bed temperatures (Martin et al. 2005). This enriching of primary air with oxygen also improves the general effectiveness of the combustion. Other positive impacts are the decreases of the flue gas flowrate and the emissions and the increase in thermal efficiency brought about by less flue gas loss. However, the consumption of electricity required for oxygen production cancels out the increase in the thermal efficiency.

6.4.2.1 Design of the Grate Firing System

The decisive parameter for the design of grate furnaces is the heating value. For residual waste, the heating value determination includes the ash and water contents. Theoretically, according to Fig. 6.29, MSW can be burned without back-up firing if the low heating value lies above 3,500 kJ/kg. However, the state-of-the-art combustion conditions required by today's environmental regulations for emissions and

Fig. 6.29 Heating value, moisture and ash content triangle (Bilitewski et al. 2000)



ash quality make it necessary to have a heating value of above 6 MJ/kg (without any additional fossil fuel). For the combustion of untreated MSW in Germany, an average LHV of 9–10 MJ/kg, with fluctuations between 8 and 12 MJ/kg, is taken as a basis.

Grate incineration is very flexible in regard to the heating value and the waste characteristics. Figure 6.30 provides an example of a waste incineration diagram, indicating the relationship between waste throughput, LHV and thermal output. Waste incinerators are designed for large variations of the average LHV, normally between 6 and 12 MJ/kg. Wastes with low LHVs require air preheating. Because of variations in the waste composition, spare steam generator capacity is required for peaks in the heat input.

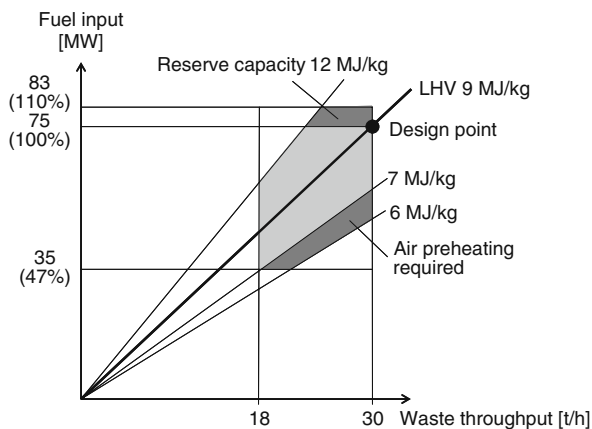


Fig. 6.30 Thermal power and throughput diagram

Typical grates are between 7 and 10 m long. The width of the grate determines the incineration capacity and can vary from 1 to 14 m for the largest units, which have a capacity of up to 40 t/h. Typical cross-sectional heat releases from the grate are in the range of 0.7–1.1 MW of fuel input per square metre (Effenberger 2000). The heat release for EFW grates is often characterised by the thermal width load, which relates the fuel input to the width of the grate. Typical width loads are 25–35 GJ/m h (related to the heat of the fuel) or 2.5–3.5 t/m h (related to the mass).

The specification of the adiabatic temperature is an important criterion in designing a waste incineration plant. Since the heating value of the waste is given, the adiabatic combustion temperature can be controlled by the air ratio or by recirculating flue gas. A low air ratio leads to high adiabatic combustion temperatures. This diminishes the flue gas volume, with the benefit that the following flue gas passes and the flue gas cleanup train can be built more compactly. A negative side effect of the higher flue gas temperatures is the shortening of the refractory lifetime. It is sometimes also suspected that a negative fouling and corrosion impact may occur (Foster et al. 2007).

6.4.2.2 Grate Variants

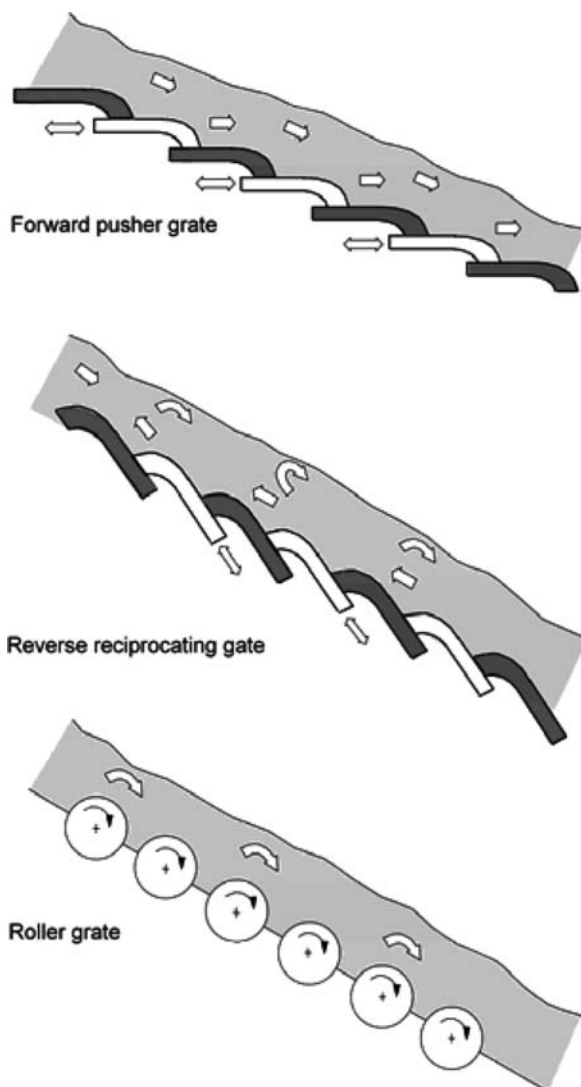
Modern grate furnaces use various grate systems, as shown in Fig. 6.31.

Pusher-type grate: The grate rods of pusher-type grates overlap and alternate between being fixed and being moveable, meaning there is a fixed rod overlapping a moving rod which overlaps a fixed rod and so on. Lifting movements of the movable rods set in motion the waste charge towards the bottom ash removal and also rake the fuel. Since the transport of the waste is brought about by the movement of the grate, it is possible to install the grate horizontally. A slight incline, though, helps the transport of the charge.

Reciprocating grates: The reciprocating grate, like the pusher-type grate, is comprised of movable and fixed grate rods, though they are arranged in a manner such that the lifting movement is directed towards the fuel inlet. The advantage of the reciprocating grate is an intensive thorough mixing of the waste charge as the rod movement constantly pushes embers from the main combustion zone back to the upper end of the grate. For this reason, the reciprocating grate is also suited to moist low calorific waste, which needs to be agitated more intensively during combustion. In contrast to the pusher-type grate which, due to the less intensive mixing, develops a prolonged drying, devolatilisation and ignition zone, the reciprocating grate features a long burnout zone. Since the inclination of the grate has to ensure the transport of the fuel towards the discharge chute, for reciprocating grates a steeper incline is required.

Roller-type grates: The roller grate is composed of rollers roughly 1.5 m in diameter arranged one after the other. The roller grate is inclined in the direction of the discharge at about 30°. The rollers are driven electrically and separately, which makes it possible to adapt the rate of feeding to the combustibility of the waste. Some agitating of the waste bulk is necessary in the gussets, the gaps between two rollers. The air is distributed via slots cut all over the surface of the rollers

Fig. 6.31 Different grate types



and through the gussets. One advantage of the roller grate is the slight amount of wear and tear, a consequence of the grate surfaces only being exposed to the high furnace temperatures briefly during the turning movement and being cooled by the primary air the rest of the time. The disadvantages are the great accumulation of “riddlings” and the fact that there is only slight grate agitation, meaning a lesser ability to intensify the combustion process.

In normal grate furnaces, the lifetime of the grate material is limited, especially the material in the incineration zone. Higher heating values further reduce the lifetime, even if specific advanced alloys are used for the grates. In consequence, water-

cooled grates have been developed. By water cooling, the temperatures can be kept significantly lower and the air distribution can be made more flexibly. However, water cooling is costly and has a negative impact on the efficiency, because at the low temperature of the cooling water the heat cannot be used in the thermal process. Approximately up to 5% of the thermal capacity is lost and is therefore not available for steam and electricity production. This corresponds to a decrease in the electrical efficiency of about 1%. With air-cooled grates, the grate design also has an impact on the grate temperatures. The reciprocating grate, owing to the intensive mixing of the grate bed and reliable covering of the grate with waste and ash, is able to manage even higher heating values without water cooling. The use of water-cooled grates, on the whole, is rare. Excessively high grate rod temperatures can be expected when the grate is not covered with fuel. This is avoided by appropriate operation.

6.4.2.3 Furnace and Boiler

The waste heat of the hot flue gases is recovered in a steam generator. EfW steam generators differ significantly from coal-fuelled steam generators, for example, by having comparatively large combustion chambers which ensure the necessary burnout and allow the substantial cooling of the flue gases. Different sections of the steam generator are distinguished:

- The first flue gas pass is the combustion chamber (or furnace) above the grate, including the primary and secondary combustion zones as well as the remaining first radiation pass. The combustion chamber size is limited by water/steam-cooled membrane wall heating surfaces, which are refractory-lined or protected by SiC tiles in the lower part. This serves to protect the metal heating surfaces from corrosion, but it also aids in reducing the heat transfer so that the stipulated flue gas temperature of 850°C can be met. In this refractory-protected area, one or several fossil fuel-fired burners are installed in the side walls. These burners are used for start-up and shutdown of the plant in order to maintain the minimum temperature conditions at all times. The membrane wall of the upper part of the furnace, from 8 to 12 m above the grate, is generally unprotected or equipped with overlay-welded nickel-based alloys (cladding). In the first part of the 2000s, a few plants with combustion chambers where the membrane walls are completely clad down to the grate have been constructed.
- Subsequently, in the second flue gas pass, there are metal, not refractory-lined, membrane-wall heating surfaces. In many EfW plants, up to three empty flue gas passes (including the combustion chamber) are built in to cool the flue gases down to low temperatures by radiant heat transfer.
- Convective heating surfaces for superheating and, in the cooler area, for feed-water heating (the economiser) are arranged, preferably in the area of flue gas temperatures below 650°C.

The design of the combustion chamber above the grate (the incineration chamber) has to be suited to the grate and depends on the chosen flue gas routing above it.

In the combustion of waste, due to the high volatile matter content, only a small part of the combustion process takes place on the grate, the larger part occurring in the area above. Therefore, it is necessary to have a sufficient residence time at high temperatures in that upper zone. The furnace geometry has to be chosen such that a high and even flue gas velocity above the grate will develop. The combustion chamber walls and the ceiling in the part above the grate are refractory-lined in order to limit the heat transfer and to protect the walls from corrosion. The refractory lining of the combustion chamber ceiling favours the heating of the fuel by radiant heat from the hot walls.

The design of the grate and combustion chamber differs according to the suppliers and their specific experiences. In general, three basic designs can be distinguished. The nomenclature comes from the flow direction of the flue gases in relation to the waste flow: co-current, counter-current and medium current, as shown in Fig. 6.32.

Co-current or parallel flow furnace: In a co-current combustion arrangement, primary combustion air and waste are guided together in a co-current flow through the combustion chamber. Accordingly, the flue gas outlet is located at the end of the grate. Only a comparatively low amount of energy is exchanged between the combustion gases and the waste on the grate. The advantage of co-current designs is that the hot flue gas passes above the ash bed and improves the burnout of the ash, which is why it is most commonly used for roller grate systems. To facilitate ignition for wastes with very low heating values, the primary air must be preheated.

Counter-current furnace: In this case, the primary combustion air and the waste are guided in opposing directions in a counter-current flow arrangement through the combustion chamber and the flue gas outlet is located at the front end of the grate. The hot flue gases facilitate drying and ignition of the waste. Special attention must be paid to avoid the slip of unburned gas streams.

Medium current furnace: The medium current design is a compromise system suitable for a wide range of fuel properties. In this case, the flue gas outlet is located in the middle of the grate. Mixture-promoting chamber contours and/or secondary air injections can be used to promote a thorough mixing of the various flue gas

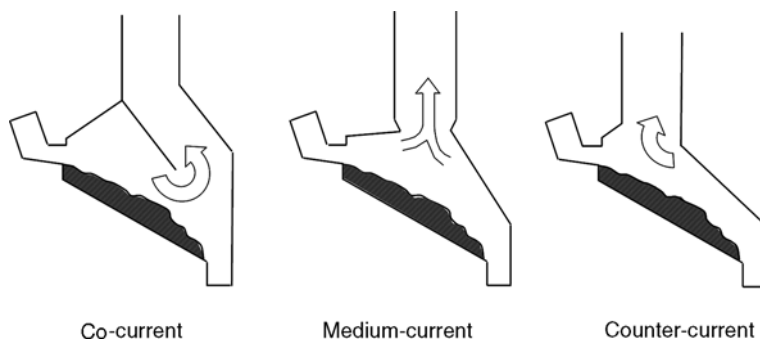


Fig. 6.32 Furnace and grate arrangements for EfW systems

streams. Nearly all modern EfW plants in Europe, the USA and Japan built in the 2000s have been designed using characteristics lying in between medium-current and counter-current furnace designs.

The primary combustion zone is followed by the secondary combustion zone, where the secondary air (or overfire air) is added for the combustion of the unburned gaseous components. The following secondary combustion conditions are favourable for a complete burnout of the flue gases:

- Availability of sufficient oxygen. An oxygen concentration of approximately 6% was considered as a minimum in the German regulation “17. Bundesimmissionschutzverordnung (17. BImSchV – the 17th Amendment to the Federal German Emissions Control Act)” of 1993. In later revisions of the 17. BImSchV (17. BImSchV 2009) this requirement was removed in order to be consistent with the EU Waste Incineration Directive (EU 2000).
- Sufficient residence time at a high temperature. According to the European Waste Incineration Directive (EU 2000), a minimum residence time of 2 s with temperatures above 850°C is required to ensure the destruction of dioxins (see Sect. 6.4.8)
- Complete mixing of secondary air and homogeneity of flue gas flow. This condition is included in the combustion engineering principle 3T, meaning that time, temperature and turbulence are required for good quality combustion.

In some cases, the primary and secondary combustion chambers are separated by a constriction. Adding the remaining secondary air prior to but near the constriction provides an intensive mixing of the flue gases, which is followed by a homogeneous flow profile on the other side of the constriction.

6.4.2.4 Ash Deposition

Even though the dust load in the flue gases of grate-based EfW plants is much lower than in fluidised bed systems, special care must be taken to avoid dust-related slagging, fouling and corrosion. Reduced underfire air flows and flue gas velocities in the furnace generally favour less carryover of fly ash to the boiler. MSW ashes contain various components such as relatively inert aluminosilicates, lime-related compounds (in the form of oxide carbonates or sulphates) and salts. The majority of the salts are alkali and earth alkali chlorides, but significant amounts of heavy metal chlorides also occur. Alkali chlorides, and especially the lead and zinc chlorides, form eutectic mixtures with low melting temperatures. Therefore, the above-mentioned flue gas temperatures of 850°C in the first flue gas path should not be markedly exceeded, because at flue gas temperatures of 1,000°C or so, sticky particles may lead to severe fouling and slagging in this section. Because it is nearly impossible to completely avoid the formation of deposits in the flue gas path, adequate measures have to be taken to control the growth of deposits and to avoid unscheduled shutdowns of the power plant for the manual cleaning of deposits. Several techniques have been developed to reduce the formation of deposits during operation. Cleaning can be accomplished with water cannons or shower cleaning, soot blowers using steam or air, mechanical rappers or explosives.

6.4.2.5 Corrosion

High-temperature corrosion is the biggest operational problem in the thermal treatment of waste, and it limits the efficiency of the conversion into electrical power. Corrosion is worst when there is, at the same time, high chlorine and low sulphur contents in the MSW and a high content of heavy metal chlorides in the fly ash. The membrane wall heating surfaces and the superheater surfaces can be severely corroded. The corrosion mechanisms are discussed in Sect. 5.10.4.

In the area of the membrane wall, corrosion is caused by molten salts. In order to diminish the effects of corrosion in the furnace above the refractory-protected area, coatings made of nickel–chromium alloys are applied by deposition welding (also called cladding) to protect the steel of the membrane wall tubes. This cladding may cover the entire upper furnace (first pass) and the first part of the second pass and is considered state of the art for the most common 40–60 bar boilers used in Europe, Japan and the USA. It can be applied either in the workshop during the construction of the plant or retrofitted if corrosion problems occur. Consequently, membrane wall corrosion no longer presents a frequent problem in EfW plants in service today.

On the other hand, however, high-temperature corrosion of superheater tubes is still an unsolved problem and limits the allowable live steam temperatures to values of 400–450°C. In this corrosion mechanism, sulphation of alkali chlorides in boiler tube deposits sets a chlorine cycle going which involves very strong wear and tear. In the design of EfW plants, this is the reason the allowable working medium temperatures are chosen as a function of additional parameters such as the flue gas temperature or the flue gas velocity, i.e. in order to avoid superheater corrosion. This design practice is represented in Fig. 6.33 by the so-called Flingern corrosion diagram.

The investigation of corrosion processes and the development of corrective measures are the subject of extensive research (Born 2007; Born 2005; VDI 2006, 2007). Corrective measures for reducing superheater corrosion take two forms: developments of new materials and combustion engineering measures. Making use of the

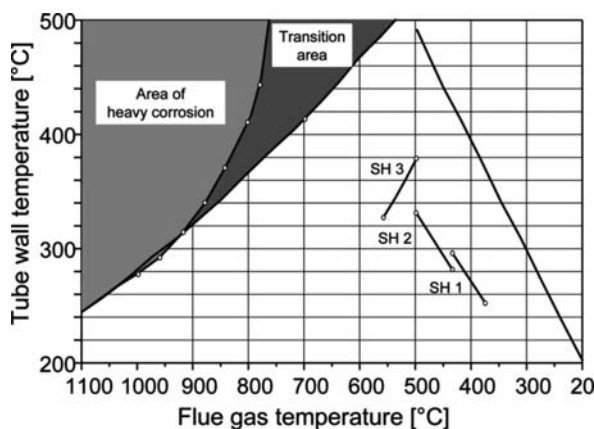


Fig. 6.33 Corrosion diagram

most highly alloyed materials and using cladding in the area of the superheater have not yet gained acceptance due to technical and economic reasons. Combustion engineering measures aim at preventing ash deposits on the superheaters or altering their consistency. For example, the sulphation of the chlorine in the fly ash before hitting the deposits can be favoured by injecting sulphur-containing additives. Another measure employed for this purpose is the use of intensive mixing and low flue gas flow velocities, which provide more time for natural sulphation. Furthermore, some authors state that by avoiding or removing deposits it is possible to diminish the corrosion potential (Warnecke 2007; Warnecke 2006; VDI 2006, 2007).

6.4.3 Pyrolysis and Gasification Systems

In the USA and Europe, interest in gasification technologies for waste processing was kindled during the 1970s, when the oil shocks struck. However, the development of the technology was unsatisfactory at that time. The European market only began to seriously reconsider the technology during the early 1990s, and this was driven by the political desire to avoid the use of incineration and to maximize recycling and resource recovery in a more sustainable way. A number of high profile companies developed processes for waste treatment that combined pyrolysis, gasification and combustion in various configurations. Most of the processes required extensive pre-treatment of the MSW (making RDF or SRF (secondary recovered fuel)).

In Japan, gasification technologies for MSW reached the highest degree of development in the late 1990s. As of 2007, more than 90 gasification plants were in operation or under construction. The main purpose of these processes is to improve ash quality and to reduce the total dioxin output per tonne of waste. Table 6.9 shows the installed capacity of gasification systems in Japan in 2008.

6.4.3.1 Pyrolysis in Rotary Kilns

The pyrolysis or gasification system which reached the highest degree of development in Europe was the Siemens SBA process (“Schwel-Brenn Anlage”) – a schematic of which is shown in Fig. 6.34. Development started in 1988 at a small-

Table 6.9 Installed capacity (in 2008) of the processes for the pyrolysis or gasification of waste realised in Japan in the 2000s (Themelis 2007)

Gasification system	Installed capacity (t/day)
JFE – thermoselect	1,980
JFE – fluidised bed	1,300
Nippon steel – shaft furnace	6,200
Ebara – fluidised bed	1,700
Other fluidised bed processes	3,200

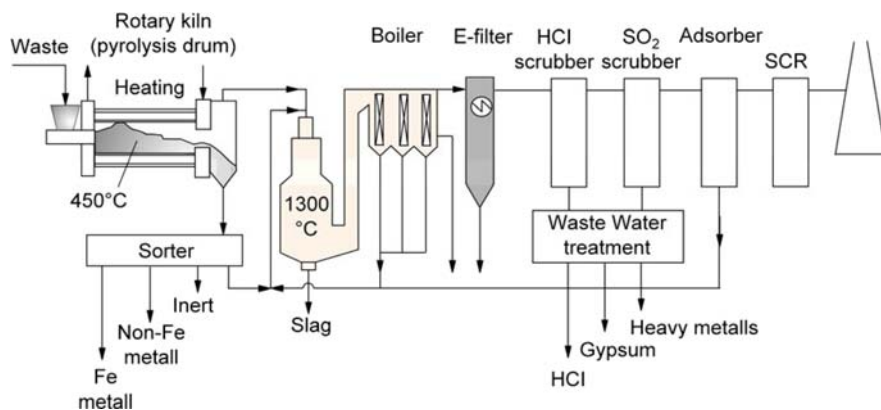


Fig. 6.34 Siemens SBA gasification of MSW (pyrolysis in rotary kiln followed by slag-tap combustion)

scale plant in Ulm, consisting of a rotary pyrolysis reactor followed by a mechanical sorting system of the residues, and then the combustion of the separated coke with the pyrolysis gases in a melting furnace. The start-up of the first large-scale plant was in 1997 in Fürth (Germany), where power generation was via a classical steam cycle. This plant was designed for a 100,000 t/year (two lines each with a 5 t/h capacity) waste throughput and had a capital cost of approximately 150 million euros. The gasification system consisted of two lines each with a 5 t/h capacity (after pre-sorting of the MSW). The SBA process had severe problems with the waste pre-processing, clogging of the rotary kiln-type pyrolysis drums and instabilities in the post-combustion melting chamber. The final decision for dismantling the plant was taken after a minor incident involving some entangled metal on 12 August 1998, which caused pyrolysis gas to escape through a damaged gliding ring seal (Schwarzmann 1999). Since then, no further projects with this technology were seriously considered in Europe, but licenses were acquired in Japan by Mitsui and Takuma. These companies constructed several commercial size plants.

6.4.3.2 Gasification with Pure Oxygen and Integrated Melting

Another gasification technology with quite a high degree of development is the thermoselect gasification and melting process, developed in Switzerland between 1985 and 1992. A demonstration facility with a capacity of 110 t/day was built in Fondotoche, northern Italy, and used to validate the technology; the facility operated under a commercial licence from 1992 until 1999 (Themelis 2007).

A larger commercial facility with a design capacity of 792 t/day (or 2,250,000 t/year) was built in Karlsruhe in Germany and commenced operation in 1999. The plant suffered technical and commercial problems and none of the lines operated at full design capacity for sustained periods (Whiting and Schwager 2006). It was finally shut down in 2004 by the owner EnBW, a large German power utility (partly

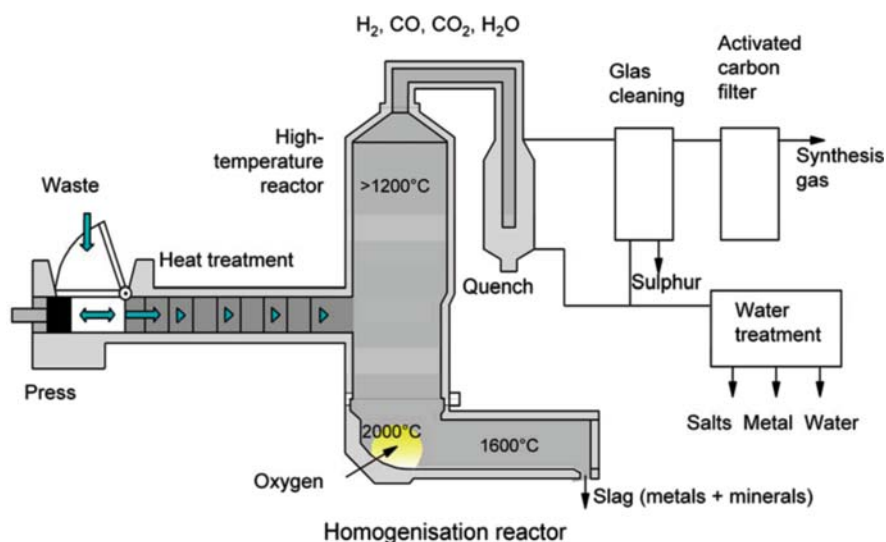


Fig. 6.35 Thermoselect gasification of MSW (gasification with pure oxygen and integrated melting of the ash as well as post combustion in a boiler)

owned by EDF – Electricité de France). Since then it has been “mothballed” pending the outcome of litigation between the supplier Thermoselect S.A. and the owner EnBW. The third European thermoselect plant, which was constructed in Ansbach, never went in operation. The schematic of the thermoselect process is shown in Fig. 6.35.

In Japan, similar plants have been built by JFE (Japan Steel Engineering), licensing the thermoselect technology. The first plant was completed in 1999 at a steel mill in Chiba with the synthesis gas produced being used in the mill. A further six JFE plants using the thermoselect technology had begun operation by 2006.

6.4.3.3 Fluidised Bed Gasification

In Europe and the USA, no highly developed fluidised bed gasification systems have been realised. In contrast, by the beginning of the 2000s in Japan, 13 companies were engaged in the development of fluidised bed gasification systems for MSW. The most utilised system in Japan is the Twin Internally Circulating Fluidised Bed Gasifier (TIFG), developed by the Ebara Corporation (Fujimura et al. 2001). It gasifies the wastes first and uses the heat content of the gases to raise the temperature in the following slag combustion furnace stage.

For all the gasification technologies that have been developed to a significant extent and have some market impact, the pyrolysis and gasification steps are only a pre-treatment. These steps are directly followed by combustion of the gases and, in the case of pyrolysis, of the pyrolysis char as well. The only exception is the fixed bed gasification technology “Sekundärrohstoff-Verwertungszentrum

Schwarze Pumpe" (SVZ) that was developed in Schwarze Pumpe/Spremberg (Brandenburg/Germany), based on German Democratic Republic (GDR – i.e. the former East Germany) coal gasification technology and used at the end of 1990s and beginning of the 2000s for MSW. Insolvency in 2004 meant the final withdrawal of this technology from the waste business in 2007 (Mielke 2007). The example of SVZ-Schwarze Pumpe seems to show that it makes little sense to use MSW, with its difficult properties, as a raw material for complex biomass to liquid (BtL) or methanol production processes, as long as coal is available.

The main driving forces for the development of gasification processes have been a desire to improve ash qualities and reduce dioxin outputs. However, the more complex gasification processes result in lower efficiencies, lower process availabilities and higher costs. Although in principle, gasification offers a better efficiency, the real plant efficiencies are even lower than conventional EfW plants. Gasification with oxygen and integrated melting has produced electrical efficiencies well below 10%, for example. Because of this, gasification technologies are no longer developed and demonstrated in Europe, while in Japan, the focus of development has been readjusted to grate-based processes producing improved ash qualities (e.g. by using oxygen enrichment, bottom ash washing, reduction of dioxin from the fly ash and the use of bottom ash in the cement industry).

6.4.4 Refuse-Derived Fuel (RDF)

The basic idea of the combustion of refuse-derived fuel by itself is that the EfW plant can be designed especially for the quality of this more homogeneous fuel. In doing so, there are design and operational advantages, and in addition it becomes possible to decentralise the plants. RDF contains chlorine, as MSW does, which means that the steam parameters are restricted in a similar way to about 60 bar and 450°C. In this respect, then, there is no advantage over waste incineration plants. However, the construction and the design of the flue gas cleaning facilities is often simpler than for EfW plants using unprocessed MSW (as a rule reduced to an SNCR process followed by calcium hydrate and activated carbon addition upstream of a fabric filter), because there are fewer pollutants. Designing the flue gas cleaning process for a narrower fuel range, though, can give rise to problems if the composition of the RDF fluctuates, especially if the chlorine and sulphur contents change (Neukirchen 2008).

The majority of RDF plants use grate furnaces. The technology is rather similar to conventional EfW plants, which have been thoroughly discussed in Sect. 6.4.2. Attention has to be paid to the higher LHV, for example, by using water-cooled grates. Fluidised beds are a reasonable technology for RDF combustion, owing to the more homogeneous fuel quality, but they are less common than grate combustion.

Four fluidised bed combustion plants were built in France between 1995 and 1997 in Mulhouse, Giens, Guerville and Monthyon (ActuEnvironment 2005). After that, no other fluidised bed projects were constructed in France. A similar scenario occurred in Japan, where 44 fluidised bed plants for MSW or RDF were

operating in 1999, with a further 11 being installed between 1999 and 2003. After 2003, there is no report of any other fluidised bed combustion plants for MSW or RDF being constructed, but various types of fluidised bed gasification processes were developed (see Sect. 6.4.3). The reasons for the low acceptance seems to be the higher pre-processing requirement in comparison to grate combustion and some negative experiences with increased dioxin formation rates in the past. The latter, however, can be solved by an appropriate design of the flue gas train, as discussed in Sect. 6.4.8.

In Germany, some circulating fluidised bed (CFB) plants were built in the 2000s to treat the increasing amounts of coarse fractions from mechanical–biological treatment (MBT) plants. One relatively well-documented example is the CFB plant in Premnitz, operated by the large German utility EON-Energy from waste. It was designed for steam parameters of 97 bar and 500°C, with superheating realised in an external bed heat exchanger. Superheating steam in the external fluidised bed heat exchanger of the recirculation loop in principle provides the opportunity to superheat steam to higher temperatures than superheaters situated directly in the flue gas path, because of a less corrosive environment. However, erosion and corrosion problems still seem to occur, resulting in a reduction of the steam temperature and pressure to 75 bar and 450°C. The CO and dioxin formation rates are much lower than in the Japanese and French examples of fluidised bed combustion and are well below the emission limits (Borghart 2008).

In most cases, RDF plants in the USA do the pre-processing of the waste on-site and utilise suspension combustion, as shown in Fig. 6.36 (Themelis 2007). RDF is

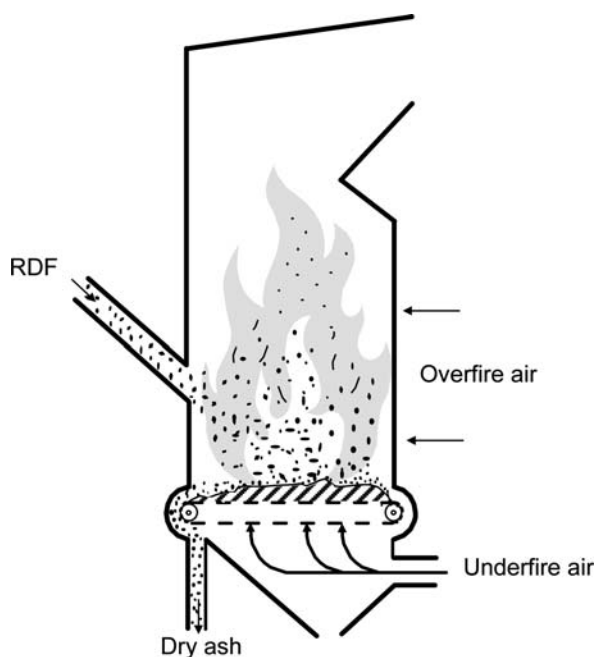


Fig. 6.36 A suspension combustion system for RDF in the USA

fed through multiple feeders into the furnace and burned part in suspension and part on a travelling grate stoker.

6.4.5 Sewage Sludge

The exclusive purpose of the monofuel combustion of sewage sludge is volume reduction for the subsequent disposal of the sludge. Power, in general, is not produced for more than in-plant use.

To ensure economic operation, the combustion of sewage sludge should develop and continue self-sufficiently without employing a support fuel. This is possible only when the heating value is larger than the power necessary to vaporise the water, heat the combustion air to the combustion temperature and compensate for the heat losses. For self-sufficient combustion, a heating value around 4 MJ/kg is required, which corresponds to a total solid (TS) content of about 40% in digested sewage sludge.

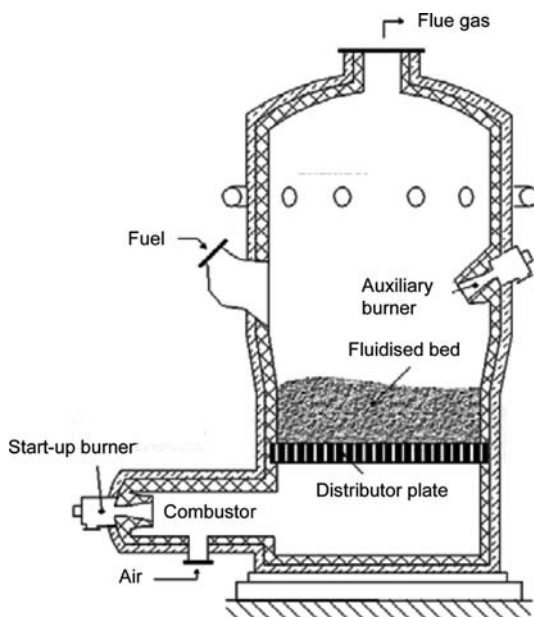
Mechanical dewatering of the sewage sludge before combustion serves to raise the heating value and to reduce the sludge volume. The degree of dewatering depends on the dewatering method and can be increased by adding conditioning agents. The total solid content after dewatering ranges between 30 and 45% TS. This is usually not enough for self-sufficient combustion. However, using a thermal drying system, it is possible to achieve TS contents of up to 95%, as Sects. 2.2.1.4 and 2.2.3.4 explain.

According to data from the German Federal Environmental Agency (*Umweltbundesamt*) in Berlin, 15 public monofuel combustion plants for sewage sludge were in service in Germany in 2004. In addition, there are six more in-house combustion plants in the chemical industry. The public and the industrial plants together had an installed capacity of about 780,000 t TS/year. The dominating technology in this context is the stationary fluidised bed. Furthermore, only three multistage grates and one single-stage swirler were in service in 2004 (Hermann 2004).

Figure 6.37 shows a sectional view of the layout of a stationary fluidised bed furnace used for the combustion of sewage sludge with a TS content of 50% (Treiber and Schroth 1992). The fluidised bed temperature has to be kept below the ash deformation temperature in order to prevent a sintering or fusion of the bed ash. The minimum temperature stipulated by law for the combustion of wastes (850°C according to *17th BImSchV*) confines the allowable operating temperatures to a lower range. The operating temperature can be set by controlling the excess air and, if needed, by heat extraction from the fluidised bed.

The good heat and mass transfer in the fluidised bed allows nearly complete combustion at a low temperature with an even temperature distribution. CO and NO_x emissions are a function of the temperature, with CO emissions decreasing and NO_x emissions increasing with a rising temperature. Temperature peaks, which are crucial for NO_x emissions, can be avoided in fluidised beds. Emission limits can be complied with in most cases.

Fig. 6.37 Bubbling fluidised bed for sewage sludge combustion (Treiber and Schroth 1992)



Another advantage of fluidised bed combustion is the possibility of capturing SO_2 in the fluidised bed. Owing to the alkaline earth matter contained in the sewage sludge, part of the sulphur dioxide is bound without any additive being supplied. For a higher degree of capture, calcium-based additives such as limestone are usually used. The capture capacity shows a maximum at temperatures around 900°C , a fact which can be used if an operational mode is sought to optimise the capture of SO_2 . However, the option of running waste incineration processes without any secondary desulphurisation measures cannot be inferred from this capture, because low SO_2 emission limits cannot be met with these operating parameters alone.

6.4.6 Steam Boilers

The heat released in the combustion of waste is used in a boiler (steam generator) to produce steam which in turn works in a steam turbine to produce electrical power. The flue gases, initially at about $1,000\text{--}1,200^\circ\text{C}$, are cooled in the boiler to temperatures typically in the range of $140\text{--}300^\circ\text{C}$. This temperature range is usually required for the subsequent flue gas cleaning process. Boilers for waste need a design suited to the often particular and rather difficult composition of the flue gas, with its corrosion and fouling effects.

The envelopment of the furnace, the following empty passes and the passes where evaporator and superheater tube bundles are located are generally designed as water-cooled membrane walls. The first pass generally needs to be empty, as hot gases are too corrosive and particulate matter is too sticky for convective heat exchangers.

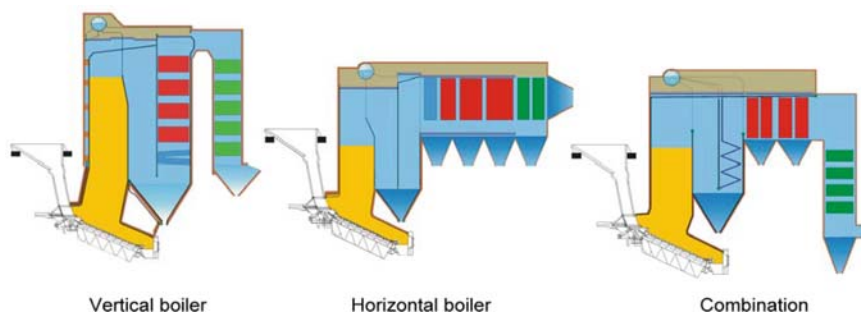


Fig. 6.38 Boiler arrangements for waste combustion (Source: Martin)

Convective heat exchangers are typically arranged in the third and following boiler passes at temperatures below 650°C . Different boiler configurations can be used in waste incineration plants. The arrangement of the heat exchangers is shown in Fig. 6.38 for a vertical boiler, a horizontal boiler (also called a “tail-end” boiler) and a combination of both. Other widely used boiler arrangements are two-drum boilers with platen superheaters in the second pass and boilers with two empty passes and a low horizontal “tail-end” section (e.g. Amsterdam).

The design of the boiler mainly depends on the flue gas characteristics (the corrosion, erosion and fouling potentials), which are themselves highly dependent upon the waste content. Hazardous wastes, for example, tend to have very wide variations in composition and, at times, very high concentrations of corrosive substances (e.g. chlorides) in the raw gas. This has a significant impact on the possible energy recovery techniques that may be employed. In particular, the boiler can suffer significant corrosion and steam pressures may need to be reduced with such wastes.

A compromise is required when setting steam parameters for waste-fired boilers. Higher steam parameters can lead to significantly increased corrosion problems, especially for the superheater surfaces and the evaporator. In EfW plants, it is common to use 40 bar and 400°C if electricity is produced, although higher values are used, especially where incentives for renewable power production are in place. In these cases, values of 60 bar and 520°C are often employed, with special measures to prevent corrosion. Because of the rather low steam parameters (low compared to coal-fired power stations), natural circulation steam boilers are selected almost exclusively. For heat production, steam at lower conditions may be produced.

6.4.7 Efficiency Increases in EfW Plants

In this section, possible improvements to waste incineration plants in service today (and how to implement them technically) will be discussed. As a reference case, a power plant with the following characteristics, produced by today’s widely implemented state-of-the-art technology, will be used:

- Steam parameters of 40 bar/380°C
- A boiler outlet flue gas temperature of 209°C
- An excess air ratio of 1.75 (flue gas O₂ content of 8.4%, dry)
- Condenser pressure of 150 mbar
- An in-plant power consumption of 2.1 MW (0.1 MWh/t of waste)

The power plant, with its live steam conditions of 40 bar and 380°C, has a net electrical power production efficiency of 20.6%. In comparison, the average efficiency of plants in service in Europe lies at about 13%.

Within the framework of a study, the potential of technically feasible solutions to improve the efficiency has been investigated using this reference power plant. The results are compiled in Table 6.10 (Gohlke and Spliethoff 2007; Spliethoff et al. 2008).

Excess air ratio: Since the development of EfW plants in the 1960s, the excess air rates have remained relatively high, between 1.8 and 2.2. This was necessary in order to compensate for fluctuating combustion conditions and to avoid the added wear from refractory and membrane wall corrosion. Today, some EfW plants have reduced the excess air ratio in conjunction with special precautions.

Mixed municipal waste is relatively homogenous and therefore well suited to decreasing the excess air ratio. Should pre-processed waste fractions or commercial waste be used, it is preferable to use higher excess air ratios (leading to lower efficiencies). The decrease of the air ratio from 1.75 for the reference case to 1.4 results in a net increase of the efficiency of 0.6% (Fig. 6.39).

Boiler exit temperature: Typically, the flue gas is cooled down in the economisers to approximately 200°C at the outlet of the boiler (209°C in the example of the reference case). The remaining energy is lost by quenching the flue gas with water to typical scrubber temperatures of 150°C or less. Alternatively, cooling can be realised by additional heat exchangers used for condensate preheating (in this case where no direct use of heat is considered). As temperatures fall below the dew points, advanced materials are needed for these heat exchangers. The reduction of the flue gas temperature to 135.5°C leads to an efficiency increase of 0.7–21.3% (Fig. 6.40).

Condensation pressure: The condenser temperature has a strong influence on the plant efficiency. The reference plant employs air condensers with a condensation temperature of 54°C, corresponding to a pressure of 150 mbar. The efficiency can be increased by 2.8 percentage points to 23.4% if a water-cooled condenser with a temperature of 23°C and 30 mbar pressure is used. However, this option for increasing the efficiency is limited in the “real world” because cooling water is rarely available (Fig. 6.41).

Steam pressure and temperature: An increase in the pressure and temperature of the steam results in an increase in the efficiency of the thermal cycle. The gain in efficiency by increasing the live steam pressure, the live steam temperature and reheating can be evaluated from a temperature/entropy ($T - s$) diagram (see Fig. 6.42). Basically, the efficiency of the thermal cycle is $\eta = 1 - (T_{\text{out}}/T_{\text{in}})$, where T_{in} and T_{out} are the average medium temperatures of heat addition and heat extraction. Steam parameters can be increased in convective superheaters up to around 73 bar

Table 6.10 Overview of measures to increase efficiencies of electricity generation (*R1* criterion of European Draft Waste Framework Directive is 0.6 and 0.65 after 2009) (Gohlke and Spliethoff 2007). D = Germany, I = Italy, NL = Netherlands, E = Spain

Measure of energy increase	Net efficiency in % (power)	Simplified <i>R1</i> criteria (only waste and power generation)		Advantages	Disadvantages	Existing plant
Typical <i>E_{FW}</i> excess air 1.75 40 bar/380°C, <i>p_c</i> = 150 mbar, <i>T_{flue gas}</i> = 209°C	20.6	0.64	0.64	High availability, moderate cost	Limited efficiency	Zella Mehlis (D)
Reduced excess air rate to 1.4	21.3	0.66	0.66	Reduced flue gas losses. Smaller equipment	Increased risk of corrosion and refractory wear in furnace. Risk of CO peaks	Brescia (I)
Wall superheater with 60 bar/460°C	23.2	0.71	0.71	Superheaters behind rear-ventilated tiles are protected by sealing air	Tiles must be removed for maintenance of superheater tubes	No large-scale reference yet
Reduced condensate pressure to 30 mbar	23.4	0.72	0.72	Reduced losses of thermal cycle. Reduced in-plant power consumption	No water cooling available at most sites. Water content in steam after turbine. Increased steam temperature or re-heat necessary	Amsterdam (NL)

Table 6.10 (continued)

Measure of energy increase	Net efficiency in % (power)	Simplified R1 criteria (only waste and power generation)	Advantages	Disadvantages	Existing plant
Increased steam parameters to 74 bar/480°C	24.0	0.73	Reduced losses of thermal cycle	Increased risk of superheater and membrane wall corrosion. Limited options for improved materials	Brescia (I)
130 bar/440°C with intermediate reheat (AEB Amsterdam energy concept)	28.1	0.84	Efficiency gain without increase of superheating temperature	Additional heat exchanger necessary (saturated steam/steam from first stage of turbine)	Amsterdam (NL)
100 bar/540°C with external superheating in combined cycle power plant	42 (referred to waste and natural gas)	–	Efficiency gain with limited corrosion risk of superheaters	Increased risk of membrane wall corrosion. High capital cost. Consumption of natural gas	Bilbao (E)

Fig. 6.39 Influence of the excess air rate on efficiency (Gohlke and Spliethoff 2007)

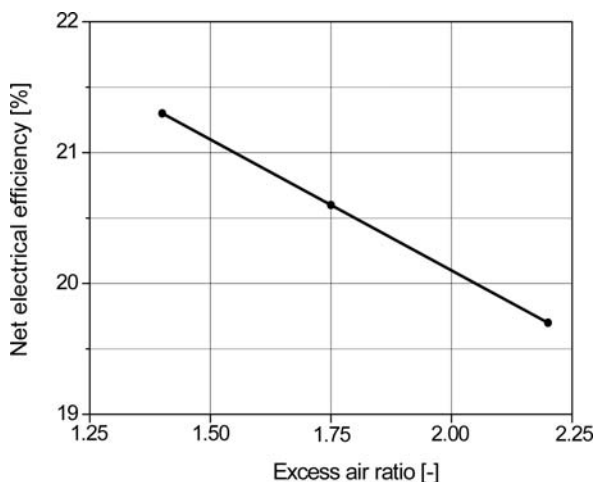
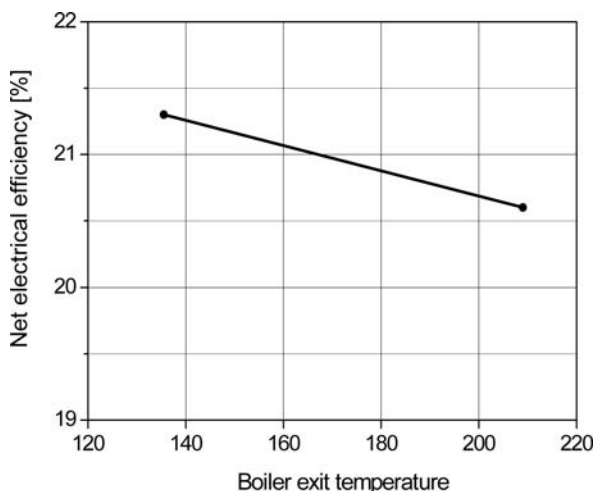


Fig. 6.40 Influence of boiler exit temperature on net electrical efficiency (Gohlke and Spliethoff 2007)



and 480°C with a classical boiler design, resulting in an efficiency increase of 3.4 percentage points to 24%. However, this increase is associated with a significantly higher risk of corrosion in comparison to a typical process which uses 40 bar and 380°C.

Another innovative approach to increase steam parameters is to use wall superheaters, where the most critical superheating tubes are placed behind rear-ventilated furnace tiles. In this way, the superheating temperature can be increased without additional corrosion problems. The critical tubes are protected by the sealing air of the rear-ventilated tiles. It is possible to obtain an efficiency increase of 2.6 percentage points to 23.2% with wall superheaters and steam parameters of 460°C and 60 bar.

Fig. 6.41 Influence of condensation pressure on net electrical efficiency (Gohlke and Spliethoff 2007)

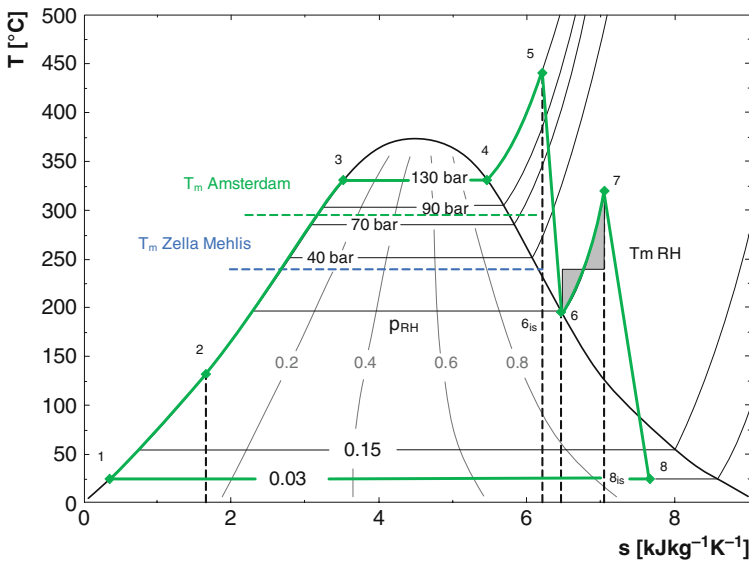
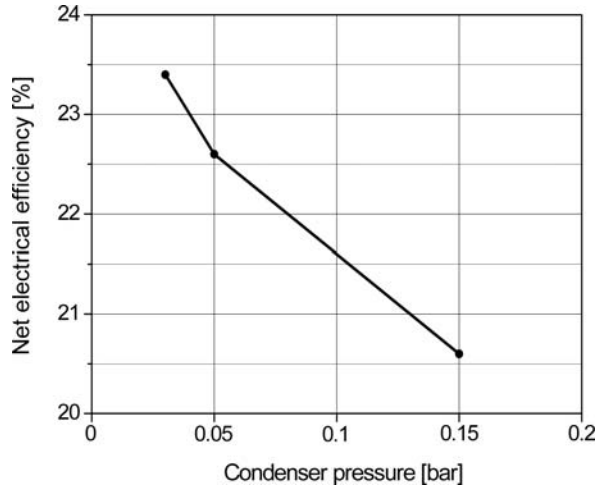


Fig. 6.42 Medium temperature of heat addition of the reference plant and of a plant with reheating (Gohlke and Spliethoff 2007)

Intermediate reheating: To achieve a further increase in efficiency, it is necessary to use intermediate steam reheating. This process has been developed by *Afval Energie Bedrijf (AEB) Amsterdam* (“the City of Amsterdam Waste and Energy Company”) and is now used in the new Amsterdam plant, which began operation in spring 2007 (Van Berlo 2006). The system is run with a steam pressure of 130 bar, a superheating temperature of 440°C and reheating of the steam, using saturated drum steam, to 320°C after the first stage of the turbine (see Fig. 6.43). Compared to the reference EfW plant mentioned above, the configuration provides an efficiency increase of 7.5 percentage points to 28.1%. In Amsterdam, an even higher efficiency

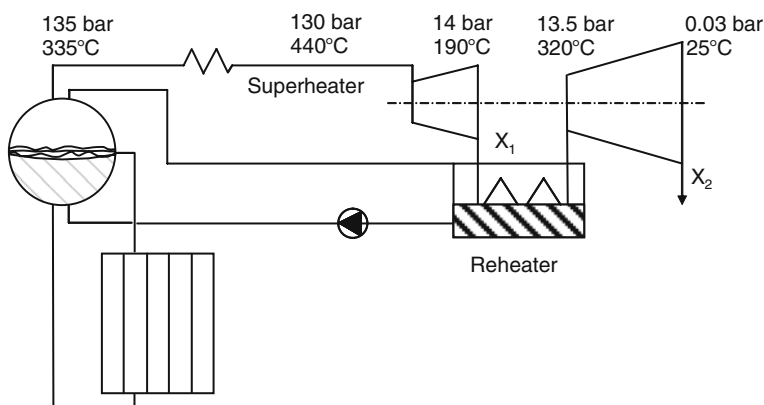


Fig. 6.43 Water-steam schematic diagram of a 130 bar/440°C system with intermediate reheating (Gohlke and Spliethoff 2007)

of over 30% is obtained, because other measures, such as reduced excess air and a lower condensate pressure, are employed as well. Due to the low reheat steam conditions at the Amsterdam plant, the higher efficiency is not the result of intermediate reheating – it is caused by the high live steam pressure; reheating is required to limit the exhaust moisture in the turbine.

External superheating: Even higher efficiencies can be obtained if superheating of the steam is performed in external fossil-fired boilers which do not have the corrosion limitations of waste-fired boilers. An example is the EfW plant in Mainz, where steam at 40 bar and 400°C is used in the intermediate reheater of an adjacent combined cycle natural gas power plant. Another example is the EfW plant in Bilbao. This waste boiler is operated with 100 bar pressure; superheating to 540°C takes place in the boiler of an integrated combined cycle power plant. In this way, the overall plant efficiency is increased to 42% (Seguin 2004).

Criterion for energy efficiency: The draft of the European Waste Framework Directive defines the efficiency criterion $R1$ as needing to be above 0.65 (from 2009) for an EfW process in order to be considered recovery (EU 2008). In the simplified case, where an EfW process does not use any additional (fossil) fuels or energy and produces only electricity, the calculation is

$$R1 = \frac{E_p \times 2.6}{0.97 \times E_w}$$

E_p : the energy annually produced and used in the form of electricity (GJ/a)

E_w : the energy supplied annually by the waste (GJ/a)

0.97: a factor taking the inevitable energy losses through radiation and bottom ash into account

The $R1$ criterion refers to the electricity produced (gross).

A typical EfW process as described above was calculated to have an efficiency $\eta_{\text{net}} = 20.6\%$ and $\eta_{\text{gross}} = 24\%$, which results in $R1 = 0.64$. In real cases, additional fossil fuels used for start-up and shutdown, as well as limited availability of the power generating components (i.e. failures, etc.), will reduce the $R1$. On the other hand, the use of heat will increase it significantly. Generally speaking, it can be estimated that EfW plants with a design similar to the reference EfW process will in fact reach the $R1$ efficiency criterion currently set at 0.6 for plants in operation and permitted before 2009 (EU 2008).

Conclusions: electrical efficiency: The main objectives of typical EfW processes are the transformation of waste into ash (incineration), the destruction of pollutants and the conversion of energy (to heat and power). A typical new EfW plant in Germany with steam parameters of 40 bar and 380°C was calculated to have a net electrical efficiency of 20.6%. This corresponds to an $R1$ of 0.64. Additional measures would therefore be necessary to meet the criterion of the European Draft Waste Framework Directive of $R1 > 0.65$ for plants in operation and permitted from 2009.

Major increases of the energy efficiency can be obtained by the following measures:

- Increasing the steam parameters (the pressure and temperature of the superheated steam)
- Reducing flue gas heat losses (via the temperature at the boiler outlet and the excess air ratio)
- Improving the steam condensation conditions (using water instead of air condensers)
- Optimising the thermal cycles (by using intermediate superheating or external superheating)
- Reducing in-plant power consumption (by using SNCR instead of SCR or a smaller excess air ratio)

Examples of recent innovative EfW plants applying these measures to increase the efficiency of electricity generation can be found in Brescia, Amsterdam, Mainz and Bilbao (Bonomo 1998; Van Berlo and de Waart 2008; Fischer 2005; Seguin 2004). The average net efficiency of electrical power generation by EfW processes is 13% in Europe (EIPPCB 2005). This could be increased in the Brescia plant to more than 25% through increased steam parameters, reduced flue gas losses and minimised auxiliary power consumption. The new plant in Amsterdam has achieved 30% with additional reheating and water condensers. Any further increase in energy efficiency is then only possible by external superheating with natural gas in combined cycle plants, as in Bilbao.

EfW is an important additional source for renewable energy, as half of the energy contained in municipal waste is biogenic. Over 50 TWh per year of renewable electricity could be generated in the EU, which is more than 10% of today's total renewable electricity generation (in the EU). To achieve this potential, it will be necessary

to avoid disposal of municipal waste by imposing landfill taxes and to put in place incentives for increasing the efficiency of EfW systems.

Combined use of heat and power: In addition to the above-mentioned measures for increasing the electrical efficiency, the combined use of heat and power should be considered. This is particularly true for the use of energy from MSW, as this fuel is usually generated close to the centres of heat consumption in the densely populated and industrial areas of the world (big cities in western Europe, the east and west coasts of the USA, Japan and China, for instance).

For plants producing both electricity and heat, it is an ongoing discussion how to weight the two products. Within the EU Waste Framework Directive, power is weighted by the factor 2.6 and heat by the factor 1.1. In Fig. 6.44, the *energy efficiency performance indicators* of the EU Waste Framework Directive and the Swiss Electricity Generation Directive (EU 2008; Bundesamt 2008) are plotted as a function of the gross electrical efficiency and the heat recovery rate. The line labelled “EU Directive $R1 = 0.65$ ” marks the minimum requirement a plant must fulfil to get the recovery status. The line labelled Switzerland marks the energy efficiency threshold for plants in Switzerland.

Figure 6.44 also includes data points derived from applying the energy efficiency criteria to different modern plants which reach particularly high efficiencies of power generation or combined heat and power generation, as well as two instances of the reference plant with combined heat and power generation.

- The example of the reference case (40 bar, 380°C) with an electrical efficiency of 20.6% achieves an $R1$ of 0.64 (if only electricity is produced) which is close to the threshold. Extracting heat from this plant reduces the electrical power production, but increases the $R1$ efficiency criterion. This is shown for heat production at temperatures of 130 and 95°C. The slope of these lines is determined by the power loss coefficient, giving the ratio of lost electricity production to heat pro-

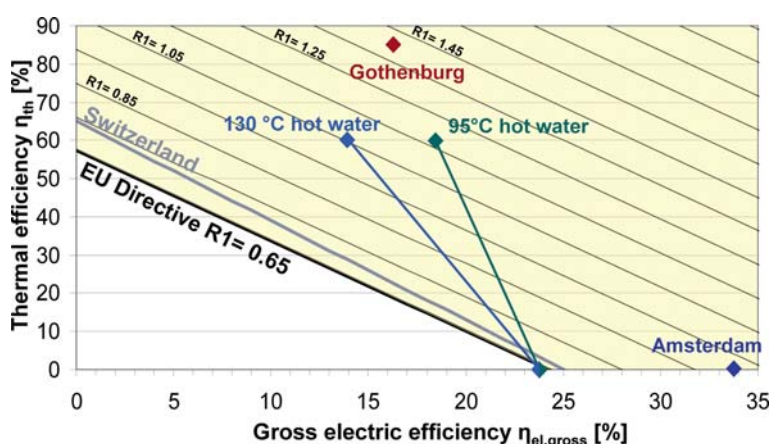


Fig. 6.44 Gross electric efficiency–heat recovery rate diagram (Gohlke and Murer 2009)

duction. The exergy efficiency, which includes the exergy of power (which is 1) and the exergy of the heat according to the temperature, is for all cases about 20.6%.

- A sophisticated heat production system in Gothenburg results in an $R1$ value of 1.42, which exceeds the required $R1$ value of 0.65 for new plants by far, whereas the exergy efficiency is only 29.7%.
- The new plant in Amsterdam focuses on the production of electricity and reaches an exergy efficiency of 30.6%, whereas the resulting $R1$ value is 0.91, which is quite low compared to Gothenburg, but still fulfils the $R1$ requirement easily.

Amsterdam and Gothenburg are two extraordinary examples with special economic and geographical constraints. However, with a reasonable combined heat and power production the $R1$ criterion can be fulfilled by state-of-the-art plants with 40 bar/380°C boilers, though for locations without a demand for heat energy, the fulfilment of the $R1$ criterion requires measures such as those described earlier to increase the electrical efficiency. The comparison of the energy performance indicators for the different plants shows that heat production is overrated if the thermodynamic value (exergy content) of heat is considered. For the author, the exergetic efficiency seems to be a more appropriate performance indicator.

6.4.8 Dioxins

The discovery of polychlorinated dibenzo-dioxins and polychlorinated dibenzofurans (PCDD/F), often simply called “dioxins”, in the flue gas of waste incineration plants had a major influence on the technical development of EfW plants. In order to point out the possibilities of dioxin reduction, the formation and destruction of dioxins will be briefly discussed in this section. The most toxic dioxin, 2,3,7,8-tetrachlorodibenzo-*p*-dioxin (TCDD), became well known as a result of the accident at the chemical plant in Seveso, Italy, in 1976. The toxic equivalent unit (TEQ) measures all toxic dioxins and furans in terms of 2,3,7,8-TCDD.

Dioxin emission from EfW plants principally results from two sources:

- dioxins may exist in the waste or they
- can be newly formed (*de novo*) when cooling down the flue gas.

Dioxins fed with the waste into the EfW plant can be effectively destroyed at high temperatures and sufficient residence time. Accordingly, a residence time of 2 s at 850°C is required in the flue gas path of an EfW plant.

The *de novo* formation of dioxins is a heterogeneous gas–solid reaction, in which the fly ash or solid carbon provides the surface for the reaction. *De novo* dioxin formation requires the presence of chlorine, oxygen and aromatic species. Gaseous HCl in the flue gas can be converted to molecular chlorine by the Deacon reaction:



Molecular chlorine reacts with aromatic species and soot in the flue gas to form PCDDs and PCDFs, the amount and particular species depending on the temperature and the boiler design. Copper in the fuel or the ashes may act as catalyst for formation. It has been shown that the de novo reaction takes place in a temperature window of 180–450°C, with a maximum formation at about 300°C, and is dependent on the residence time of the gas and in particular the fly ash in that temperature range.

Effective primary measures to reduce the de novo formation of dioxins are as follows:

- A complete burnout, which reduces the potential of dioxide formation by destroying the aromatic compounds and soot. The residence of 2 s at 850°C, which serves to destroy existing dioxins also promotes a complete burnout.
- A low residence time of the flue gas and fly ash in the temperature range of 180–450°C. This can be achieved by rapid cooling or quenching of the flue gases. Particle filters should be installed at lower temperatures, preferably below 180°C.

Dioxin emissions can be effectively reduced by primary measures to destroy existing dioxins in the fuel and to prevent de novo formation by appropriate boiler design and flue gas handling. Dioxin emissions are influenced by the burnout behaviour and depend on the temperature course of the flue gas train. High dioxin emissions from EfW plants in the past were mainly caused by de novo formation in particle filters installed at excessively high temperatures in combination with an incomplete burnout. Additionally, dioxins can be reduced by secondary flue gas cleaning. Secondary measures can be catalytic dioxin reduction or adsorption on char coal, which then has to be treated thermally. As a result of emission control requirements, incineration in developed countries is now a very minor contributor to dioxin emissions (Baumbach 1990; Nussbaumer 2004).

6.4.9 Flue Gas Cleaning

To comply with the stipulated emission limits of the European Waste Incineration Directive, a great number of methods for each individual pollutant are available. In the flue gas cleaning process, the following pollutants have to be removed from the flue gas:

- Fly ash (removed by ESP or bag filter)
- Sour gases like HCl, HF and SO₂ (removed by wet scrubbing, spray dry absorption, etc.)
- Heavy metals like Hg, Pb, Zn (removed by activated carbon filter or filtration)
- Organic substances like PCDD/F (removed by activated carbon filter or filtration)
- Nitrogen oxides (removed by SNCR or SCR)

Since the different methods of flue gas treatment have largely been discussed in Chap. 5, they shall be examined in the following only in regard to their use in waste incineration plants (Nethe 2008; Vehlow 2006).

Dust separation: Both fabric filters (also called baghouse filters) and electrostatic precipitators (ESPs) are suitable for removing particulate pollutants. To combine dust separation with wet absorption of acid gases (HCl , SO_2), ESPs are usually employed. They are cheaper and associated with a lower pressure loss than baghouse filters. The removal of dust and pollutants is lower with ESPs, but sufficient if associated with wet scrubbers for the additional removal of pollutants. In combination with spray dryer processes, the fabric filter is advantageous, because the sorptive effect on the filter linings can be exploited. The operating temperatures of baghouse filters are usually between 140 and 200°C. ESPs are also used at temperatures up to 280°C, upstream of the last economiser (last in relation to the flue gas stream; first in relation to the condensate water stream) in combination with SCR catalysts.

Separation of acid gases: For the separation of acid flue gas components such as HCl , HF , SO_2 and SO_3 , wet and dry processes rival each other. Wet processes use scrubbing slurries, putting them into close contact with the flue gas to be cleaned. The separation process usually runs in two steps: in the first step, HF , HCl and Hg compounds are scrubbed with water and in the second step, SO_2 and SO_3 are separated by the addition of a lime slurry or sodium hydroxide. Wet scrubbing methods, with their good mass transfer between the gas and the liquid, are very effective and work under nearly stoichiometric conditions, so the consumption of absorbents is low. A drawback is the wastewater that is produced, which is of a type requiring wastewater treatment or concentration by evaporation. Wet processes are preferred in countries where it is authorised to discharge effluent to the river, such as in Switzerland and Austria or to the sea like in the Netherlands.

Dry or spray drying processes, in contrast, do not produce wastewater. In dry absorption or in an entrained-flow absorber, solid absorbents like calcium hydrate or sodium carbonate are fed to the reactor to separate the acid components, whereas in spray drying, an aqueous lime slurry is finely atomised and completely evaporated. The good mass transfer between the gas and the liquid in spray drying is again advantageous. The salt particles formed as a consequence of evaporation of the water and chemisorption are removed from the gas flow in a filtering separator. In this case, fabric filters offer the advantage of further removal via the solid layer of matter on the filter. Spray drying processes typically run at temperatures of 150–170°C.

A further reduction of organic pollutants or heavy metals can be effected by adding surface-active adsorbents such as activated carbon or activated charcoal. Dry or spray drying flue gas cleaning systems are preferred in countries where it is difficult to get authorisation for effluent discharge to rivers or seas. This is the case in most western and southern European countries as well as in the USA and Japan. In Germany, dry or spray drying processes are favoured because of the option of disposing the fly ash and flue gas cleaning residues in abandoned salt mines at

relatively low cost. In France and Japan, these residues are solidified with cement or concrete, which is costly, and there is contention over the long-term stability of such procedures.

Toxic heavy metals (e.g. Hg, Cd, Pb, Zn) and organic substances (e.g. PCDD/F): The technologies used for these pollutants are entrained-flow and fixed bed adsorbers. The methods to separate heavy metals and organic matter are based on the adsorption of the pollutants by carbonaceous surfactants such as activated carbon or lignite coke and not on absorption as in the removal of acid gases. Entrained-flow processes are designed for the separation of heavy metals and dioxins by adsorption onto reactants which are injected to the flue gas stream. In a fixed bed adsorber, the separation process occurs as the flue gas flows through a packed bed of carbonaceous adsorbents. Single-chamber systems with fillings of activated carbon/lignite coke and multi-chamber systems with various adsorbents are used.

Both entrained-flow and fixed bed adsorbers can be used as safety or “police” filters at the end of the flue gas cleaning train. Remaining heavy metals or dioxins are removed by adsorption onto carbonaceous material and remaining acid components can be absorbed by the addition of calcium hydroxide.

NO_x reduction: Nitrogen oxides can be reduced by primary measures, by selective non-catalytic reduction (SNCR) or by selective catalytic reduction (SCR). SCR technology not only gives the highest reduction rates but also involves the highest cost. In the low-dust SCR configurations which are most common in Europe and Japan, the catalyst is arranged after the scrubber to prevent deactivation. This configuration has the disadvantage of the need to reheat the flue gases to the operating temperature of the catalyst (above 240°C). In the USA it is common to use SNCR (and not SCR) in EfW plants.

Flue gas cleaning configurations: All modern waste incineration plants are equipped with an efficient flue gas cleaning system which guarantees a reliable compliance with emission limits. Figure 6.45 shows possible process variants. Configurations with wet gas cleaning (configuration a) are clearly more complex than dry processes. In German plants with wet gas cleaning, concentration of the blow-down solutions by evaporation is required, which can be performed by an external evaporator or by installing a spray dryer and a fabric filter into the hot flue gas stream (configuration b). In contrast, the dry variant in combination with NO_x control by SNCR is clearly less complex (configuration c). By means of the spray dryer, both the acid gases and mercury and dioxins can be separated by adding activated carbon. In the following fabric filter, the products from flue gas cleaning are removed together with the fly ash. In some plants, the fly ash is removed separately beforehand.

The dry process has the disadvantage that it has to be operated with a higher sorbent stoichiometry and so produces considerably higher amounts of residues, which, because of their solubility in water, are more difficult to dispose of (Vehlow 2006). Police filters were used in Germany and Austria in the 1990s but are not common in newer plants because of the general advancement of flue gas cleaning technologies.

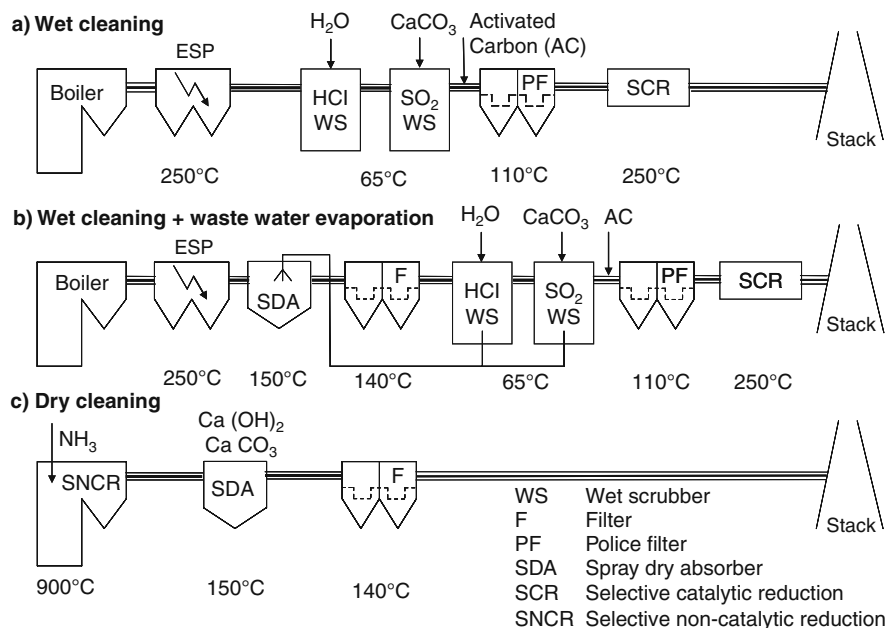


Fig. 6.45 Configurations for flue gas cleaning

6.5 Co-combustion in Coal-Fired Power Plants

The combined thermal utilisation of biomass and fossil fuels provides a cost-efficient option for the short-term exploitation of the biomass that is currently available. Given that biomass is a solid fuel, it is reasonable from the technical point of view to use it in combustion or gasification plants in combination with other solid fuels. Existing coal firing power plants, with outputs up to a maximum of 2,500 MW_{th}, are almost all combustion plants and, because of their firing and flue gas cleaning installations, are ideally suited to the co-combustion of biomass (Spliethoff 2000; Spliethoff et al. 2001).

The share of the biomass in the total thermal output of a co-fired power plant is limited by the biomass flow that can be used without inhibiting the operation of the plant and also by the quantity of biomass that can be supplied to a given plant site. This second constraint sets a limit of around 50–100 MW of bio-fuel energy input. Higher capacities strongly increase goods traffic, which becomes problematic when attempting to gain the approval of authorities. A fraction of 10–30% of biomass in the total thermal output from a pulverised hard coal plant should not be exceeded, so that adaptation measures in the plant can be kept to a minimum. In brown coal-fired and fluidised bed furnaces, it is possible to use greater fractions. The exact values have to be determined for each plant individually.

Power generation by co-firing biomass in existing coal-fired power plants has a number of advantages over generation in small plants fired with biomass only:

- The large total existing power plant capacity – even with a relatively small biomass fraction of the total fuel input – allows biomass utilisation at a large scale without a delay for construction of biomass-specific plants. This holds true despite the fact that not all large power plants are generally suitable for the co-combustion of biomass.
- The power generation efficiencies of large power plants are high compared to the efficiencies of small plants fired with biomass only.
- In case of seasonal non-availability or of shortfalls in bio-fuel supply due to weather conditions, the generation of power can still be guaranteed based on coal (i.e. a high security of supply combined with high fuel flexibility).
- The additional capital costs needed to co-fire biomass in existing coal-fired power plants are relatively low compared to new dedicated biomass combustion systems. While the capital costs for the new construction of a local biomass combustion plant amount to between 2,500 and 3,000 €/kW_{el} of installed capacity, retrofitting an existing power plant for co-combustion requires capital costs of about 300 €/kW_{el} of biomass input. In the main, these additional costs are allotted to fuel preparation (VGB 2008).

Besides biomasses such as wood and straw, other biomasses that are produced in adequate and homogeneous quantities can, in general, also be used in co-combustion, for example, sewage sludge. Co-combustion is also suitable for biomass types that are rather problematic from the combustion engineering point of view. In Germany, if the co-fired fuels derived from waste are within the legally defined range of between 0 and 25% of the thermal capacity, the plant can still be licensed and operated according to the German Ordinance on Large Combustion Plants (13th *BImSchV*). Only part of the statutory regulations for waste incinerators (re 17th *BImSchV*) apply to co-combustion (see also Sect. 5.6).

Co-firing technologies are presently being routinely commercially practiced in the USA, Finland, Denmark, Germany, Belgium, the Netherlands, Austria, Spain, Australia, Britain and a number of other countries. An inventory of the application of co-firing worldwide in 2004 indicated that more than 150 coal-fired power plants had experience with co-firing of biomass or waste, at least on a trial basis. A large range of biomass materials including herbaceous and woody materials, wastes and energy crops are co-fired today (Van Loo and Koppejan 2008; Baxter 2005; Fernando 2005; Fernando 2007).

The focus of co-combustion activities varies from country to country and depends on the availability of biomass types, but mainly on legal issues and national incentives. Many countries have initiated incentives to encourage the use of biomass for electricity production. Some examples are as follows:

- According to a political agreement in Denmark from 1993, power stations in Denmark have to use 1.4 million tonnes of biomass, of which at least 1.0 million tonnes must be straw, every year beyond 2000 (Berg and Jensen 2008).
- In 2002, the Dutch government agreed a coal covenant with the six major utilities in which they were obliged to reduce CO₂ emissions from coal by 5.8 Mt/year in the period 2008–2012. More than half of this target, namely 3.2 Mt, is to be

achieved by the substitution of biomass for coal. This equates to an installed biomass capacity of 508 MW. To achieve these targets, the incentives for electricity production from biomass of a subsidy of 4.8 € cents/kWh and a tax redemption of 2.9 € cents/kWh were put in place. In the Netherlands, all eight coal-fired power plants co-fired biomass, industrial waste, RDF or sewage sludge up to 2005 (Fernando 2005).

- In Germany, the Renewable Energy Act promotes the utilisation of biomass only in units up to 20 MW_{el}. Electricity generated from clean biomass is subsidised at the rate of 7.7–10.6 € cents/kWh (from 2009: 5.8–10.2 € cents/kWh). The lower numbers are for larger plants in the range extending to 20 MW_{el}. Due to this reason, clean biomass is hardly used for co-combustion. The focus of co-combustion is on waste fuels, which are not subsidised. In Germany, nearly 20 plants co-fired sewage sludge and another 6 co-fired industrial waste or RDF up to 2007 (Fernando 2007).

6.5.1 Co-combustion Design Concepts

The different co-combustion design concepts are shown in Fig. 6.46. Co-firing can be direct, with biomass and coal being fed into the same boiler, or indirect, where a pre-treatment like gasification is carried out prior to the combustion in the main unit. In parallel combustion, biomass and coal combustion are separate and the boilers are connected only on the steam side.

Direct co-firing of biomass in furnaces is a simple method of combined biomass and fossil fuel utilisation. The firing technologies to be considered for the co-combustion of solid biomass and waste matter are those used for coal – grate, fluidised bed and pulverised fuel firing. In direct co-firing, the additional fuel is simply added to a boiler designed for the base fuel, usually coal. This is the most convenient method, able to be used in connection with both fluidised bed and pulverised fuel boilers. Fluidised bed combustion is quite suitable for co-combustion because of its

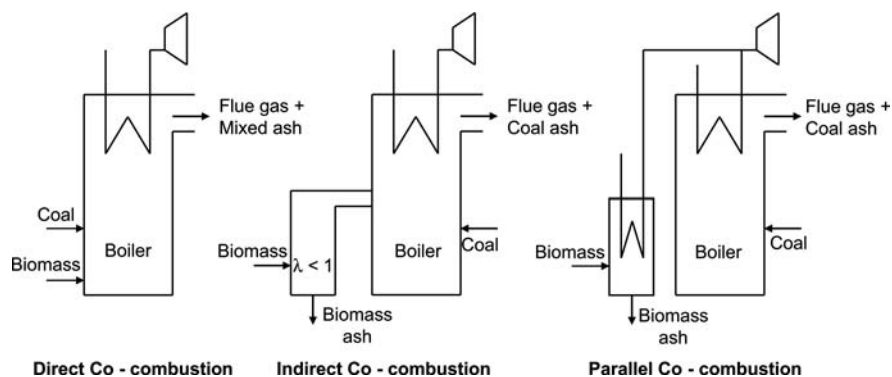


Fig. 6.46 Co-combustion arrangement options

fuel flexibility, whereas the pulverised coal combustor requires a well-defined fuel size distribution. Experience shows that only minor quantities of additional fuels (a few per cent of the fuel power) can be prepared together with the coal in the existing coal mills of a PC plant. If larger quantities are to be used, special mills and burners are desirable. There are no such limitations in FBC. In fluidised beds, larger quantities and larger fuel particle sizes can be used (Leckner 2007; Spliethoff 2000).

Because there are only a small number of grate firing systems used for coal, this technology shall not be considered further in the context of co-combustion. Grates are used for mono-combustion of biomass and predominantly for waste incineration, since their capacities are adequate for the quantities of waste produced at many locations and the technology is suitable for a wide range of fuel types.

Direct co-combustion of biomass in coal-fired plants, however, can have negative effects on operation and the quality of the residual matter. It can impair the plant performance through fouling, slagging, or corrosion, or reduce the potential for use of the fly ash. An additional pre-treatment step for the biomass such as pyrolysis, gasification or washing can solve these problems (see Fig. 6.47).

In contrast to direct co-combustion, it is possible to separately remove and utilise the ashes of coal and biomass by topping with a pyrolysis or gasification step. Combustible, low-calorific-value gas is injected and burned in the main combustor, whereas most of the ashes remain in the gasifier, from where they are removed,

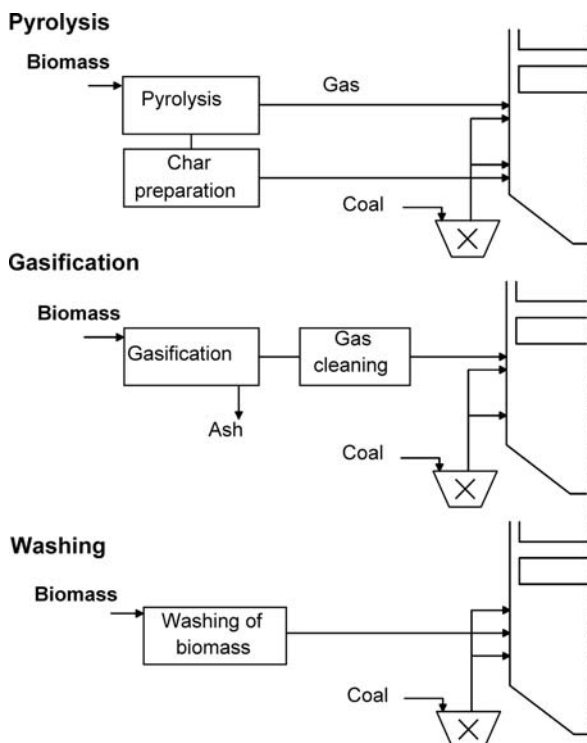


Fig. 6.47 Indirect co-combustion configurations

separately to the ash removal from the main boiler. However, gaseous components such as volatile alkali compounds or trace elements from the additional fuel will enter the main combustor if no adequate gas cleaning is provided for. Such systems have been built in the Netherlands, Finland and Austria. Used as a reburn fuel, the pyrolysis gases are able to reduce the nitrogen oxides formed in the combustion of the pulverised fuel. Fouling or erosive effects caused by biomass ashes are avoided or reduced, and the commercial value of the fly ash from the coal is not impaired (Leckner 2007; Fernando 2002).

In parallel co-combustion, biomass is fired in an entirely separate combustion system which is connected to the main boiler only on the steam side. Steam produced in the biomass combustor can be either fed directly to the joint turbine or to the coal boiler to be superheated. In this way, the additional furnace can provide heat to the steam cycle at a suitable steam temperature. Another advantage of connecting only on the steam side is that any harmful substance released in the additional boiler will not affect the main combustor, and both ashes and flue gases can be treated separately. This promotes the utilisation of both the coal and the biomass ash. An example of such a plant is the 600 MW_{el} multi-fuel power plant at Avedøre, Denmark. Straw of a thermal input of 100 MW is burnt in a grate combustor, with the steam produced having the same steam conditions as the main boiler – 300 bar and 580°C. The multi-fuel concept enables efficiencies of 48% for the main unit and 45% for the straw unit (Noppenau 2003; Fernando 2002).

Direct co-firing is straightforward but can lead to several technical problems. Indirect and parallel co-firing incurs greater costs and is in particular suitable for biomass containing troublesome or harmful components or when the quality of the ash is of importance. It is necessary to weigh up the costs and the other pros and cons of the available technologies. Indirect and parallel co-firing is less common than direct co-firing.

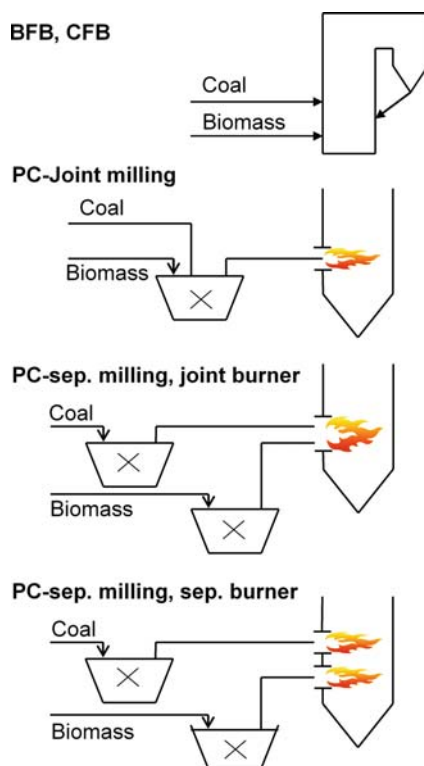
6.5.2 Biomass Preparation and Feeding

The required preparation for bio-fuel for direct co-combustion in existing power plants depends on the type of biomass and on the firing technology. For wood, woodchips can be regarded as the state of the art. Experience in Denmark suggests that providing herbaceous biomass in the form of bales is suitable.

Different methods for supplying the fuel to the furnace can be distinguished. Preparation for and feeding of PC boilers usually require more steps, for example, milling or even separate burners for the additional fuels. The methods are shown in Fig. 6.48.

The simplest option is to blend the bio-fuel with the coal, introducing the mixture through the existing fuel-handling system and through the existing pulverised coal burners. The main restriction is the different milling behaviours of coal and biomass. Most coal mill designs base the pulverisation of the coal on its brittleness. This milling principle can be used in the case of sewage sludge, but it is not

Fig. 6.48 Fuel supply arrangements for PF and FB co-firing

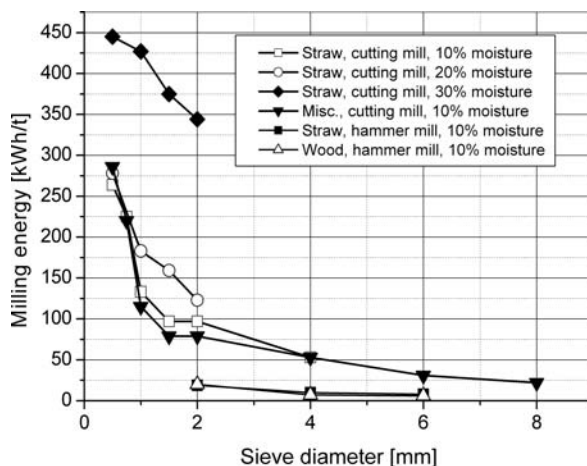


suitable for woody or herbaceous biomass. For these fuels, the technologies usually applied are cutting milling or hammer milling. If the biomass has a considerable moisture content, the heat balance of the furnace limits the addition of the fuel to a few per cent of the total fuel power. For straw and wood co-firing, separate milling in cutting or hammer mills, as well as separate feeding to the burner, is the standard technology. The pulverised fuel can be injected into the furnace together with coal at the same burner or to a separate biomass burner. For fluidised beds the system is less complicated as there are no burners or grinding devices.

Wood in the form of woodchips can be used in fluidised bed furnaces without further milling. Maximum chip sizes are about 50 mm in circulating fluidised bed furnaces and about 90 mm in stationary FB furnaces. Bales of herbaceous biomass have to be undone and the biomass cut into chaff of lengths between 10 and 30 cm. So in regard to fuel particle sizes, there is practically no difference between co-combustion and combustion with biomass only. Overall, fluidised bed firing requires very little preparation of the fuel and is in addition insensitive to hydrous fuel types with low calorific values.

To achieve complete combustion, pulverised fuel firing requires the biomass (e.g. woodchips or bales of herbaceous biomass) to be milled to a high degree, adequate devices for which are cutting and hammer mills. In order to determine the

Fig. 6.49 Milling energy required for cutting and hammer mills of different sieve insert diameters (Siegle 2000; Spliethoff 2000)

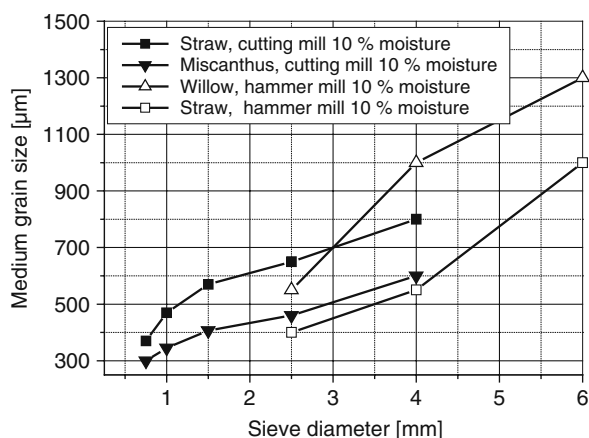


energy requirement for milling, tests were carried out using a cutting mill and a hammer mill. In both systems, the biomass fibres were repeatedly reduced in size until they fell through the holes of a sieve insert. The electrical energy needed for this size reduction increased as the particle size (i.e. the sieve insert hole size) decreased, ranging between 0.8 and 2% of the calorific value of the biomass if a cutting mill was used with sieve holes between 2 and 6 mm (see Fig. 6.49). With a hammer mill, the energy required was only 0.5–1% of the calorific value (Siegle 2000). As a rule, the biomass milling energy requirement increases with the moisture content of the biomass – with moisture contents of more than 10–20%, the energy demand rises especially high. When straw with a moisture content of 30% and a sieve hole of 2 mm was chosen, the energy demand was more than 8% of the calorific value. A cutting mill was used, however, which is less efficient for such a task.

Wood particles produced in cutting mills have a rather cubic form, whereas straw and *Miscanthus* particles are small, elongated, rectangular plates. Figure 6.50 shows the average particle diameters determined by sieving for various sieve inserts, milling methods and biomass types. It should be observed that fibres longer than the diameter of the sieve holes can fall through the sieve. For these reasons, the values in Fig. 6.50 are only approximate.

The bio-fuel has to be appropriately milled to guarantee complete combustion during pulverised fuel firing. In a 0.5 MW experimental plant, to give an example, the maximum particle size to ensure ignition and complete combustion was a diameter of about 6 mm for straw and about 4 mm for *Miscanthus* (due to its wood-like structure). Wood co-fired with pulverised coal required a milling degree of 2–4 mm (Kicherer 1996; Spliethoff and Hein 1996). In co-combustion in plants with higher thermal capacities of up to several thousand megawatts, coarser particles can also be fired because of the longer residence times. In one industrial plant, for instance, straw with stem lengths up to 10 cm could be burned completely (Bemtgen et al. 1995).

Fig. 6.50 Medium particle size as a function of sieve diameter (Siegle 2000; Spliethoff 2000)



Preparation and feeding of mechanically dewatered or (additionally) thermally dried sewage sludge can usually be achieved using the existing installations for coal handling and preparation. For this purpose the sewage sludge is added before the milling process, so that it is completely dried, milled and injected into the firing together with the coal. If, however, only mechanically dewatered sewage sludge is fed to the mill (i.e. not thermally dried sludge) along with the coal, the amount of sewage sludge for co-firing is limited by the mill's capacity to evaporate the moisture. The process of milling and drying the fuels together ensures that coal and sewage sludge are well mixed.

Another possible requirement might be to dry the biomass. While it is usually not necessary to dry wood and straw for the actual combustion, there may be an energy efficiency advantage in doing so, in particular when previously unused waste heat can be utilised. Sewage sludge, in contrast, does as a rule require at least partial drying. Mechanical dewatering by centrifuge or compartment-type filter press can raise the dry matter fraction of the sewage sludge from about 3–5% to between 20 and 45%. This process, though, is already carried out at the sewage treatment works. Combustion or thermal drying can effect a further decrease of the water content. While mechanical sludge dewatering is in all cases used at sewage treatment plants for volume reduction, the application of thermal drying is debatable. The issue, however, is not whether adequate process technology makes thermal drying superfluous, but where the drying should be carried out (external to or inside the combustion plant), how the dewatered sewage sludge should be integrated into the thermal process, what the costs are for the individual solutions and what the outcomes are of the respective energy balances.

For an external drying process, using a low-energy drying medium can optimise the heat utilisation. Sewage sludge drying at a power plant offers the opportunity of employing either low-temperature steam, flue gas or combustion air. The less favourable option is drying at the sewage sludge works either by using a dryer fired with gas collected from the treatment works (mainly methane) and/or natural gas or directly in a firing system.

6.5.3 Co-combustion in Pulverised Fuel Firing

The issues of prime importance when co-firing biomass in existing pulverised fuel firing plants are the mass and volumetric flowrates and their rates of change. They must be compared to those of coal firing alone. The existing fuel transport arrangements and the fuel preparation process must be suitable for the new volumetric fuel flow and have adequate capacity. Likewise, the change in the moist volumetric flue gas flowrate has to be determined, because of its substantial influence on the heat transfer and the residence time in the steam generator and on the functioning of the downstream flue gas cleaning equipment.

When biomass is co-combusted in existing coal-fired power plants, attention must be paid to the impacts on all the units occurring as a result of the bio-fuel properties and the deviation from the design fuel. Figure 6.51 identifies the possible impacts on the components of a pulverised coal-fired power plant. These impacts will be discussed in the following detailed descriptions of wood and herbaceous biomass co-firing. In addition, sewage sludge is taken into consideration as a supplementary fuel because, so far, it is industrially the most significant waste fuel.

6.5.3.1 Volumetric and Mass Fuel Flowrates and Flue Gas Flowrate

The volumetric flow of the fuel increases considerably as a result of the admixture of ligneous (wood and wood waste) and herbaceous biomass to the coal. This is due to the far lower calorific value and the greater fuel volume of the organic material com-

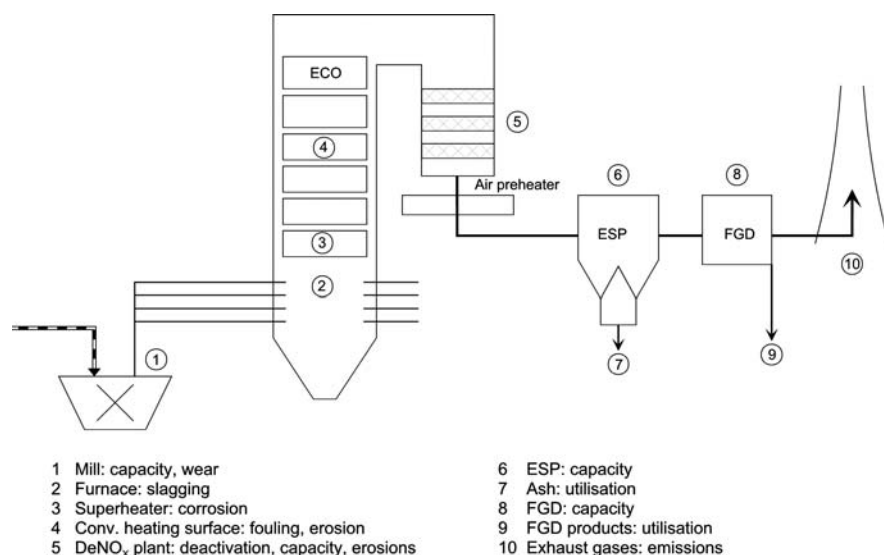
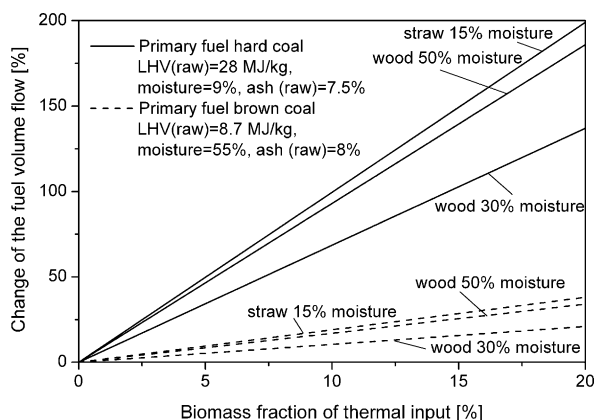


Fig. 6.51 Possible impacts of co-combustion (Spliethoff 2000)

Fig. 6.52 Increase in the volumetric as-received fuel mass flow in biomass co-combustion (bulk density of coal = 870 kg/m^3 , brown coal 740 kg/m^3 , chopped material (30% moisture content) = 250 kg/m^3 , straw bales (15% moisture content) = 150 kg/m^3)



pared to coal (Fig. 6.52). To give an example, the total volumetric fuel flow doubles when straw is co-fired in pulverised hard coal firing when straw has a 10% fraction of the thermal input. For this reason, both milling and transport to the furnace should be separate. Often there is no choice, as the coal milling installations in most cases cannot be used for biomass because of the difference in material structures. Biomass therefore requires specific milling installations.

The change in the volumetric wet flue gas flowrate depends on the moisture content of the primary fuel and the biomass. Co-firing of straw with a moisture content of 15% with hard coal as the primary fuel results in only a slight increase, of 5%, when the biomass input is 20% of the total thermal input. In contrast, very moist biomass (e.g. bark or fresh cuttings) may have a stronger effect and the resulting increase in the flue gas volume may limit the substitutable fuel quantity. If brown coal with a high moisture content is being fired as the primary fuel, the flue gas flow decreases slightly (see Fig. 6.53).

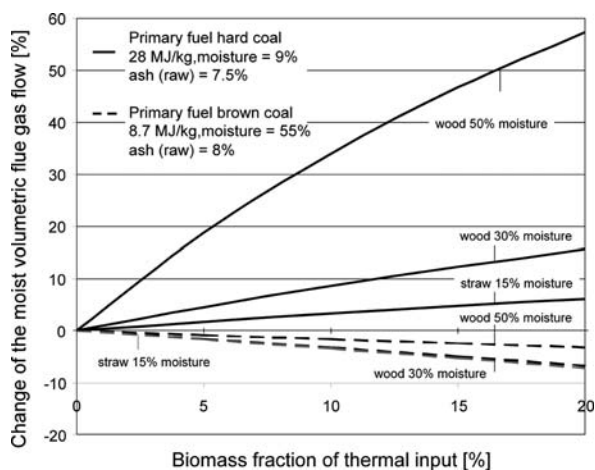


Fig. 6.53 Change of moist flue gas volume in biomass co-combustion

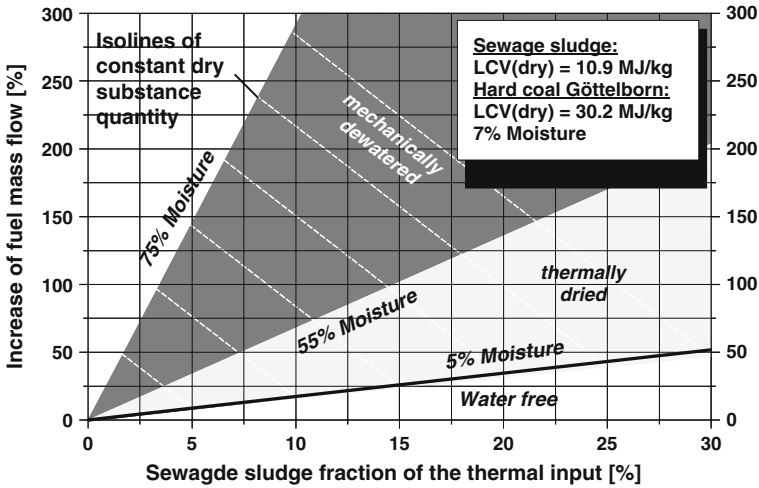


Fig. 6.54 Influence of co-combustion of sewage sludge on the fuel mass flow (Gerhardt et al. 1997)

The influence of the moisture content of sewage sludge in co-combustion with hard coal is described in Fig. 6.54 for the fuel mass flow and in Fig. 6.55 for the volumetric flue gas flow (Gerhardt et al. 1996, 1998). Both diagrams are presented as a function of the sewage sludge fraction of the fuel heat input. When reading these charts it must be remembered that the same fraction of sewage sludge with a higher moisture content would need more dry substance to be fired with it for the evaporation of its moisture.

For this reason, the diagrams in Figs. 6.54 and 6.55 also feature isolines of equal dryness and constant population equivalents (the population equivalent gives the

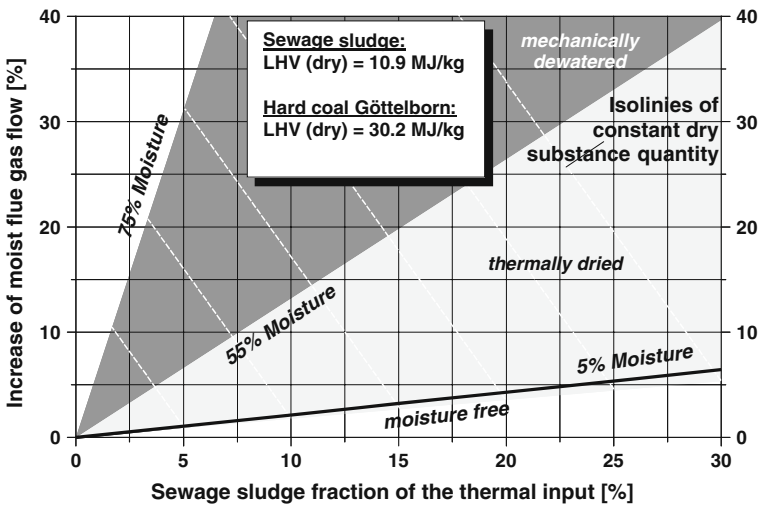


Fig. 6.55 Influence of sewage sludge co-combustion on the moist flue gas flow (Gerhardt 1997)

average amount of sewage sludge to be disposed of per inhabitant). While the fuel mass flow amounts to 18% for a thermally dried sewage sludge with 95% dryness by weight, contributing a 10% fraction to the thermal capacity, the fuel mass flow rises to about 70% for mechanically dewatered sludge at 45% dryness by weight if the same thermal fraction is to be contributed. In this latter case (mechanically dewatered sewage sludge), a higher dry substance mass flow is fed to the firing, which means that a greater sewage sludge quantity, given in population equivalents, is disposed of. If the same population equivalent is taken as a basis both for the mechanically dewatered and for the thermally dried sewage sludge, the result is a 50% larger fuel mass flow for the mechanically dewatered sludge, as seen in Fig. 6.54. The thermal input fraction is around 7% compared to 10% for thermally dried sewage sludge.

There is a strong case for the mechanical handling equipment, the mills and the fuel feeding via the burners to be modified if the sewage sludge will bring excessively high moisture into the installed equipment of the power plant. The same is true, more or less, for the flue gas system. The higher flow velocity leads to changed heat transfer rates and also impairs the separating performance of the flue gas cleaning train.

If the primary and additional fuels have nearly the same moisture contents, only minor modifications are necessary. In practice, therefore, either mechanically dewatered sewage sludge is combined with raw brown coal (moisture about 50%) or thermally dried sewage sludge is combined with hard coal (moisture about 7%). If there is only a small fraction of mechanically dewatered sewage sludge, however, it is also possible to use it in hard coal firing.

6.5.3.2 Combustion Process

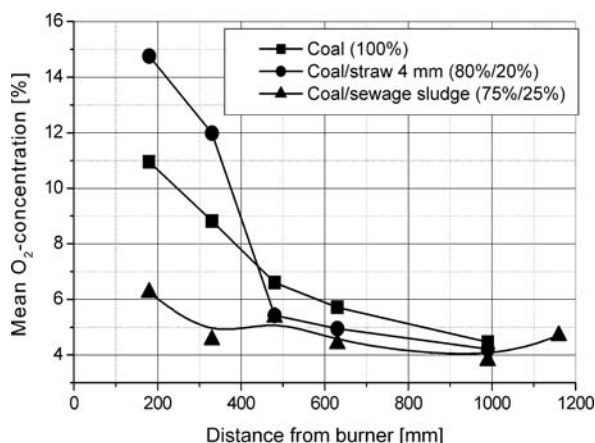
Due to the high volatile matter content of biomass in comparison to coal, much coarser biomass particles can be used in pulverised fuel firing. Herbaceous biomass types are more reactive than wood in this respect and therefore require less fine milling.

The significantly coarser milling degree of the biomass particles has an impact on the combustion process. Figure 6.56 shows the course of combustion for three fuel combinations: pure coal, 20% milled straw/80% coal and 25% sewage sludge/75% coal. The gradients of the mean oxygen concentration and the temperatures (*not shown*) of the combustion courses illustrate how the ignition of the coarse straw particles is delayed. After ignition, though, oxygen is consumed faster and the combustion proceeds more rapidly to completion in comparison to firing coal only. For thermally dried sewage sludge with a similar particle size distribution to coal, combustion proceeds faster.

6.5.3.3 Slagging, Fouling, Erosion

The process of deposit formation is described in detail in Sect. 5.10. There are two principal mechanisms through which co-combustion can affect slagging and

Fig. 6.56 Course of the combustion process of a mixed biomass/coal firing



fouling. The first mechanism is dependent on the melting of the bulk ash and is mainly related to the formation of molten deposits (Tortosa Masiá 2006). When biomass is co-fired with coal in low to medium ratios, the behaviour of the resulting ash will be dominated by the coal ash. However, the presence of significant levels of alkali and alkaline earth compounds in the mixed ash can change the behaviour of the coal ash, reducing ash fusion temperatures by 100–200°C and promoting formation of molten deposits. This effect is more pronounced for coal ashes with high fusion temperatures. In these cases co-firing at even modest ratios can have a major impact on the ash fusion behaviour. The effect is less dramatic with coal ashes originally having lower fusion temperatures and already having significant slag formation propensities. The effects of lower ash fusion temperatures differ between dry-bottom and slag-tap furnaces. While low fusion temperatures may be rather welcome in slag-tap furnaces, in dry-bottom firing they can lead to slagging in the combustion chamber, especially around the burner, thus impairing the firing process.

When co-firing woody or herbaceous biomass fuels (which have moderate ash contents), the volatilisation and condensation of alkali metals is the major mechanism for the initiation and growth of fouling deposits. Alkalis vaporise at flame temperatures, undergo chemical transformations and settle on surfaces that have suitable temperatures for condensation. The sticky layer of condensed alkalis acts as a glue for other solid ash constituents, thus initiating deposit formation. Most types of biomass or wastes are high-fouling fuels, and in almost all cases co-firing with coal increases the likelihood of fouling in comparison to coal alone. On the other hand, deposit formation is lower than in pure biomass combustion (Fernando 2007; Leckner 2007).

Although fouling and slagging problems on convective heating surfaces are increased by the addition of low-melting biomass ash, investigations at a 0.5 MW plant showed that the fouling rate in straw co-combustion is only slightly higher than in the combustion of a low fouling tendency coal. Dust layers were easily removed as well. If the biomass fraction is not very high, the coal ash characteristics

dominate. However, when the fraction is larger, incompletely burned straw particles form deposits, resulting in slagging because of the low fusion temperatures of straw ash (Heinzel 2004).

The large-scale experience in Europe suggests that slagging and fouling are unlikely to be a problem for co-firing ratios less than 10%. At one plant using a 20% co-combustion ratio of straw, the boiler performance was still satisfactory, although there was need for additional soot blowing, and some slagging occurred (Fernando 2005).

With sewage sludge, lower fusion temperatures and a higher ash content than the standard fuel may be the causes of slagging inside the furnace. The ash deformation temperatures of investigated sewage sludges have been found to be between 1,200 and 1,300°C, 100°C lower than the temperature of hard coal ash (Gerhardt 1998). In slag-tap firing, lower fusion temperatures may be favourable, but they can cause slagging in a dry-bottom furnace. The high ash content may cause erosion of the convective heating surfaces, a risk that also occurs with poultry litter, which contains a high ash content as well. However, the impacts of the much higher ash fraction of sewage sludge should still be controllable by soot blowing.

6.5.3.4 Corrosion

Herbaceous biomasses such as straw or cereals have significantly higher chlorine contents than coal and most other bio-fuels, which may increase high-temperature corrosion of the heat exchange surfaces. The high-duty surfaces of the superheater, where there are high steam and flue gas temperatures, are most likely to be affected. In a Danish 130 MW_{el} pulverised coal furnace, corrosion tests were carried out for two superheater materials. Using a 10% thermal input fraction of straw, the corrosion rates were moderate and only slightly higher than in coal mono-combustion (Fig. 6.57) (Bemtgen et al. 1995; Spliethoff and Hein 1995b). The results were con-

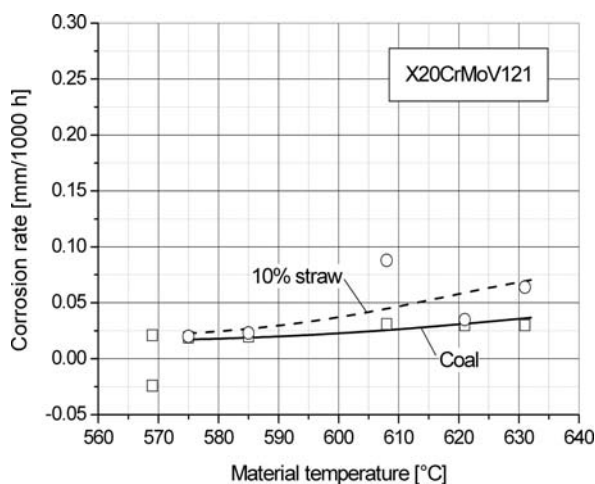


Fig. 6.57 Corrosion rates of straw co-combustion in a 130 MW_{el} pulverised fuel firing system (Spliethoff and Hein 1995; Bemtgen et al. 1995)

firmed in a 2-year investigation into co-firing with straw in a 150 MW_{el} coal-fired boiler. Up to steam temperatures of 580°C, the corrosion risks seemed tolerable. The introduction of a 20% straw fraction increased the corrosion rates by a factor of 1.5–3 at steam temperatures up to 540°C. Nevertheless, the corrosion rates are the same as those of low to medium corrosive coals (Wieck-Hansen et al. 2000). During these tests the fly ash and deposits that were formed mainly consisted of potassium aluminosilicates and potassium sulphate. Most of the chlorine left the boiler as less corrosive HCl, and there was little KCl in the deposits (Andersen et al. 2000). This example demonstrates the beneficial effect of the minerals and the sulphur in the coal on the transformation of the potassium in the straw (Leckner 2007).

Waste fuels containing high levels of alkalis or chlorine, such as refuse-derived fuel (RDF) and poultry manure, can increase corrosion rates by factors comparable to straw. RDF, if it has a high plastic content, can contain chlorine levels five to six times higher than average coals. Any efforts to increase the efficiency of the unit by increasing steam parameters will promote high-temperature corrosion. Sewage sludge has chlorine levels similar to coal and hence is less likely to cause high temperature corrosion.

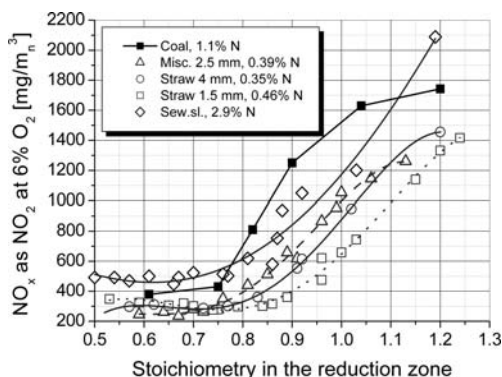
In pulverised coal co-combustion, the extent of corrosion by high alkali and chlorine fuels is reduced by the presence of sulphur dioxide in the flue gas, an effect not occurring in monofuel combustion. SO₂ reacts with alkali and alkaline earth chlorides to form less corrosive sulphates. These sulphate compounds, however, are only stable under oxidising conditions, hence corrosion may still occur in overfire air systems (Fernando 2007).

6.5.3.5 Emissions

One result of co-firing ligneous and herbaceous biomass is a reduction of the major pollutants in the cleaned flue gas – sulphur dioxide (SO₂) and nitrogen oxides (NO_x). CO emissions do not rise if the biomass has an adequately high milling degree. Supporting evidence came from investigations within the framework of an EU project, which produced more detailed information about emissions from different plant types co-firing herbaceous and woody bio-fuels (Bemtgen et al. 1995).

NO_x emissions: The high volatile matter content is the reason why biomass is especially suitable for the application of NO_x reduction measures like air and fuel staging (Spliethoff and Hein 1998). Although the fuel nitrogen in straw in relation to the calorific value is in the same order of magnitude as the content in coal, the result of the higher release of pyrolysis products and volatile nitrogen compounds from straw is less NO_x formation. Figure 6.58 shows the NO_x emissions measured at an experimental plant co-combusting different biomass types with hard coal, the biomass comprising 25% of the thermal input. The results reveal roughly the same emissions of NO_x, regardless of the fuel nitrogen content. The suitability of biomass to nitrogen-reducing combustion engineering measures is indicated by the decreasing NO_x emissions at diminishing primary air ratios. Co-combustion of biomass in pulverised coal firing therefore does not depend on the biomass nitrogen content. Even higher concentrations in sewage sludge or poultry litter can be controlled, to

Fig. 6.58 NO_x emissions with air staging for different biomass types, biomass fraction: 25% (Kicherer 1996; Spliethoff and Hein 1996)



a certain extent, by combustion engineering measures. Biomass fuels promote the formation of ammonia instead of HCN in the primary combustion zone (DiNola 2007; Di Nola et al. 2009). Detailed investigations into the emission behaviour of biomass-coal blends in different combustion processes are described in Kicherer (1996). Due to its high volatile content, biomass is also suitable for reburning to reduce NO_x emissions. As shown in Fig. 5.53, pulverised *Miscanthus* as a reburn fuel is nearly as effective as natural gas. A pyrolysis gas from biomass can even be better as a reburn fuel than natural gas (Rüdiger 1997).

Despite the much higher fuel nitrogen content of sewage sludge, the NO_x concentrations after the fuel-burning system are not correspondingly higher, as they are limited by a lower conversion rate. However, within the typical range of use, up to 25% of the thermal input, an increase in emissions when incorporating sewage sludge co-firing must be expected if there are no additional methods to reduce NO_x . By applying the investigated in-furnace reduction methods of air and fuel staging, NO_x emissions from sewage sludge co-combustion were comparable to the combustion of coal only. Still, the high fuel nitrogen content should be taken into consideration when designing and constructing the furnace and considering the distribution of air in the combustion chamber.

SO_2 emissions: In pulverised coal combustion, the sulphur contained in the fuel is almost completely oxidised into sulphur dioxide (SO_2), so that SO_2 emissions can be directly correlated with the sulphur input. Figure 6.59 plots SO_2 emissions as a function of the fraction of biomass or sewage sludge, respectively, in investigations carried out at a 0.5 MW furnace. Increasing the share of biomass (i.e. wood, straw or *Miscanthus*) caused a decrease in SO_2 emissions. This reduction can be attributed to the low sulphur input. However, it is observed that sulphur is additionally captured in the biomass ash. Starting from about 90% in pure coal combustion, the conversion rate drops linearly to reach a value of 50% in pure biomass combustion. The coarse milling of the biomass and its moisture content delay combustion, thus favouring, because of the lower temperatures, the capture of SO_2 in the biomass ash (Spliethoff and Hein 1998).

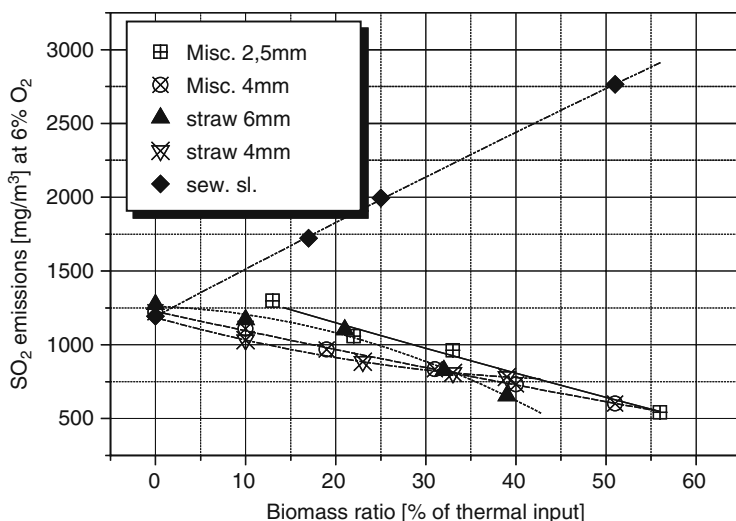
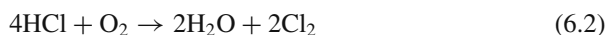


Fig. 6.59 SO₂ emissions as a function of the biomass ratio for different blends. (Kicherer 1996; Spliethoff and Hein 1996)

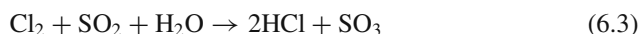
In co-combustion of thermally dried sewage sludge with hard coal, the concentrations of SO₂ and NO_x after the firing depend on the input fraction of the sewage sludge, the fraction of sulphur or nitrogen in relation to the calorific value and the conversion rate. The nitrogen contained in the sludge, in relation to the calorific value, is about —six to eight times higher than that contained in the coal. For fuel sulphur, the ratio is —three to four times higher than coal. The sulphur-to-SO₂ conversion rate of about 90% is not affected by the sewage sludge fraction, so the result is a rise in SO₂ emissions in proportion to that fraction. The high CaO content of the sewage sludge ash did not have any reducing effect on SO₂ emissions in the tests. This inactiveness of CaO can be explained by surface sintering as a result of the high combustion temperatures in pulverised fuel firing, as sewage sludge is milled to a similar degree to coal.

HCl emissions: The chlorine content of herbaceous biomass fuels such as *Miscanthus*, grass and straw and also of waste fuels such as RDF and municipal solid waste (MSW) can be considerably higher than coal. Straw can contain chlorine concentrations of about 1%, which is about 10 times greater than typical bituminous coals. The reactions of chlorine have been discussed in the context of deposit formation. In pulverised coal co-combustion, the availability of sulphur will lead to the sulphation of alkalis and formation of HCl, which is beneficial with respect to corrosion. HCl is completely removed by FGD scrubbers.

Dioxins: Fuels containing chlorine are suspected of producing harmful polychlorinated dibenzo-*p*-dioxins (PCDDs) and polychlorinated dibenzofurans (PCDFs), especially if the ashes of the fuels contain copper or other catalysts for dioxins. HCl in the flue gas can be converted to molecular chlorine by the Deacon reaction:



Molecular chlorine reacts with aromatic species in the fuel to form PCDDs and PCDFs, depending on the temperature and the boiler design. When chlorine-containing biomass and waste fuels are co-fired with coal, the formation of PCDDs and PCDFs is inhibited. The possible mechanisms for PCDD/F inhibition include the depletion of molecular chlorine concentrations by the reaction with SO_2 :



Results from laboratory and large-scale investigations show that in co-combustion, due to the SO_2 in the flue gas, the emissions of PCDD/F are as low as for coal-fired plants (Fernando 2007; Leckner 2007).

6.5.3.6 Effects on Residual Matter

The effects of biomass co-combustion in coal-fired power plants on residual matter should be split into two: the consequences of a greater ash load and the effect on the commercial exploitability of the ash.

Ash load: The low ash content of wood and straw bio-fuels reduces the workload of the dust removal equipment. All in all, less ash will be produced in comparison to coal firing alone when co-firing ligneous and herbaceous or petiolate bio-fuels. In contrast, the typically high ash contents of sewage sludge dry matter can lead to higher workloads for the electrostatic precipitator when larger sludge fractions are used.

Commercial exploitability of fly ash: The composition of the fly and bottom ashes in coal firing determines their possible uses. For the utilisation of fly ash in the cement and concrete industries, the critical parameters are the concentrations of alkalis, SO_3 , Cl, CaO and unburned carbon.

Since fly ash utilisation was one of the major obstacles for the broad application of co-combustion in hard coal-fired power plants, a review of EN 450 was initiated in 1999 and, since 2005, a new European Standard “fly ash for concrete” has replaced the earlier one. Fly ash from co-combustion of specific secondary fuels such as woodchips, straw, olive shells, cultivated biomass, municipal sewage sludge and paper sludge can now be used for concrete if the percentage of secondary fuel does not exceed 20% by mass of the total fuel, the derived amount of ash from the co-combustion material is not greater than 10% of the total ash and the requirements of the fly ash quality can be met (Wiens 2005). The maximum allowed contents of total alkalis, Cl and residual carbon for fly ash to be used in concrete production are 5% by wt., 0.1% by wt. and 5% by wt., respectively (see Sect. 5.11).

The commercial operation of straw co-firing at the 350 MW_{el} pulverised coal-fired Studstrup Unit 4 in Denmark began by burning straw with a maximum share of only about 10% on an energy basis. The fly ash is presently used for cement production (Zheng et al. 2007).

In the Netherlands, a programme was carried out to determine whether the fly ash produced from co-firing met the quality requirements for its utilisation as a filler material or as a cement replacement and for the production of concrete. Tests were undertaken at several power plants using up to 10% by mass of the secondary fuel. The fuels used included several types of pellets, sewage sludge, pet cokes, wood chips, poultry manure, MBM (meat-and-bone meal) and liquid hydrocarbons. The resulting fly ashes demonstrated that even high biomass co-firing percentages can produce fly ashes that meet European standards (Fernando 2007).

The ash from brown coal firing is commonly used as a filler material in opencast mines, and the regulations extend to this use, with the most critical parameter being the leachability of the ash components.

Waste fuels such as sewage sludge or RDF may contain higher concentrations of heavy metals. The evaluation of the ash properties, therefore, and in contrast to coal ash, above all has to take into account these heavy metals, which display an accumulation behaviour in the biosphere. Heavy metal concentrations in waste fuels are highly dependent on the origin of the fuel. As far as municipal sewage sludges are concerned, surveys certify that despite the given variations only a few sewage sludges have higher pollutant concentrations than is permitted for use in farming.

Apart from substances such as mercury, selenium and arsenic, which escape in elemental or compound form in appreciable fractions in the flue gas flow because of their low boiling points, most of the trace elements from sewage sludges are found in the solid residues from the furnace or the flue gas cleaning processes. Figure 6.60 shows a direct comparison between heavy metal ash concentrations of sewage sludge ash and of typical hard coal ash (Gerhardt et al. 1996; BMU 1996; Fahlke 1994). According to this comparison, the trace element concentrations, taking into account the different ash contents, approximate each other. The co-combustion of sewage sludge and hard coal therefore does not result in a serious change of pollutant concentrations in the ash.

For heavy metals that are partly carried out in the flue gas flow from the plant, it is necessary to check the removal efficiency in the cleaning sections downstream of

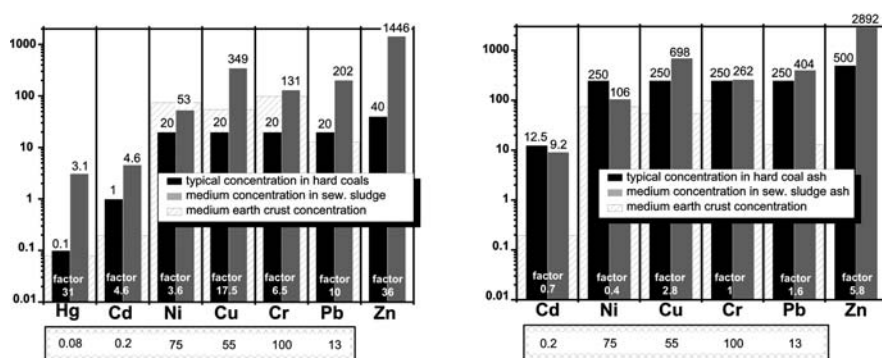


Fig. 6.60 Concentration of trace metals in dry fuels and ashes (Gerhardt et al. 1996; BMU 1996; Fahlke 1994)

the furnace. In the wet flue gas desulphurisation units common in power plants, only some of these heavy metals are removed, so the remaining part is emitted as part of the exhaust gas flow (Tauber et al. 1996). For mercury, for instance, a removal efficiency of about 50% is given (Fahlke 1994); the rest gets emitted. This circumstance can make it necessary to improve the common flue gas desulphurisation systems in power plants, for instance, by the addition of specially adapted precipitants or by the downstream installation of an additional filtering stage (e.g. an activated charcoal filter).

6.5.3.7 NO_x Control Equipment

When considering the impacts of co-firing on NO_x control, two scenarios have to be considered: plants with high-dust removal configurations and those with low-dust configurations. There is a smaller impact on low-dust configurations, because NO_x cleaning is preceding by the gas cleaning steps of ESP and FGD, which mean low-dust configurations are better suited to co-firing than high-dust ones.

Common practice in hard coal-fired dry-bottom furnaces is to install NO_x control in high-dust configurations. This puts the catalyst at risk, especially if straw is used, when employing biomass co-combustion. Various mechanisms may work to deactivate the catalyst. One of these is based on reactions of the catalyst with potassium and sodium. Accordingly, catalyst manufacturers set limits on the alkali fraction ($K_2O + Na_2O < 4\%$ by weight of ash). Depending on the coal type used, this amount may be reached even with small straw fractions. Another mechanism is triggered by alkalis and alkaline earths blocking up pores of the active catalyst cells. Arsenic and phosphorus, too, can poison the catalyst. Catalyst deactivation can be limited by installing the catalyst after the flue gas desulphurisation unit – that is, by employing a low-dust configuration. In brown coal firing, where sulphur is removed by combustion engineering techniques, this problem does not occur.

Testing of SCR catalyst elements in the slip stream of a power plant resulted in high deactivations when co-firing with 20% straw. After 3,000 h of operation the catalyst activity was reduced by 35% with a high-dust configuration, whereas with a low-dust configuration employing dry flue gas desulphurisation, the loss was between 10 and 15%. Due to the set-up of the facility, the test conditions are considered to be the worst-case scenario. More than 7,500 h can be expected before the activity is reduced to 50% (in the high-dust configuration), which is still considered a high level of activity (Wieck-Hansen et al. 2000). Results of 2 years of 7% co-firing of straw at Studstrup Unit 4 showed that there was no decrease in the removal efficiency of the high-dust SCR (Fernando 2005).

The high ash content of sewage sludge can cause fouling and erosion in high-dust configurations. Because of the high nitrogen content of sewage sludge, the NO_x concentrations after the furnace may rise to a level such that they have to be reduced by NO_x control.

In co-combustion processes with sewage sludge and meat-and-bone meal as secondary fuels in hard coal-fired power plants, an increased deactivation of the SCR-DeNO_x catalysts for flue gas denitrification was observed. Investigations revealed

a correlation between the phosphorus content in the fuel and the degree of catalyst deactivation. In combustion, the phosphorus is released from the fuel, leading to increased concentrations of both particulate and gaseous phosphorus compounds. Gaseous phosphates in particular penetrate the catalyst surface, effecting severe deactivation. Particulate phosphorus, too, contributes to the deactivation by obstructing the catalyst pores and reacting with sulphuric acid. A calcium addition can abate the deactivation by phosphorus (Beck 2007).

6.5.3.8 Flue Gas Desulphurisation (FGD) Equipment

The low sulphur content of biomass reduces the load on the flue gas desulphurisation plant. However, the increase in other flue gas components as a result of the use of the biomass may impair the function of FGD or necessitate additional FGD capacity. These potential consequences set a limit on the biomass fraction, particularly in regard to the chlorine input into the FGD.

Besides sulphur, the FGD unit also removes a number of other flue gas components. The volatile ash components leaving with the flue gas, such as mercury, arsenic, lead and other heavy metals, are partly removed together with the FGD residual matter, the quality of which has to be checked if it is to be commercially used. However, the concentrations of these substances in the biomasses that have been investigated can be neglected in comparison to coal. This is not the case when municipal sewage sludge is co-fired – in this case, the quality of the residues can sometimes be negatively affected.

Sulphur in sewage sludge exceeds the usual sulphur content of coal considerably. The FGD unit must have sufficient capacity to deal with this additional load. With a 25% thermal input fraction of sludge, the SO_2 to be dealt with rises to 1.6 times the quantity from coal alone.

6.5.4 Co-combustion in Fluidised Bed Furnaces

Fluidised bed furnaces are suited to a wide range of fuels, including biomasses such as wood or straw and wastes. Biomass can be co-fired both in bubbling and in circulating fluidised bed furnaces. Co-combustion in a fluidised bed is uncomplicated and in most cases limited only by the heat balance of the bed. When co-firing with herbaceous biomass or waste fuels, steam conditions can be limited by the need to avoid deposition and corrosion (Leckner 2007).

Given its higher volatile content, biomass tends to have post-combustion reactions in the freeboard volume of the furnace, in particular in bubbling fluidised bed furnaces. In these furnaces, lightweight particles such as straw can easily be carried away from the fluidised bed, which raises the temperature in the freeboard if they post-combust. The well-mixedness of a circulating fluidised bed creates an even distribution of furnace temperatures. Nevertheless, an upward temperature shift can also be observed.

As fluidised bed firing is especially suited to high-ash and high-moisture fuel types, it seems a good technology for mechanically dewatered sewage sludge. While

the higher moisture content of the sludge increases the volumetric flue gas flow when co-firing with hard coal, there is little effect when co-firing with brown coal.

The impact of biomass co-combustion on gaseous emissions was investigated at experimental and industrial plants of various thermal capacities within the framework of a research project funded by the European Union (Bemtgen et al. 1995). Except for HCl emissions in straw co-combustion, the results showed that biomass addition has a positive effect. In all plants, the observed result was a reduction in SO₂ emissions with an increasing biomass fraction of the thermal input. This effect can be put down on the one hand to the low sulphur contents of the biomasses and on the other hand to the fact that SO₂ is captured in the biomass ash.

The correlations between NO_x emissions and co-combustion were diverse. With a low biomass fraction, the emissions of NO_x changed very little in some of the plants. In other plants, the emissions were reduced, in particular by co-firing wood (the reduction increasing with the wood fractions).

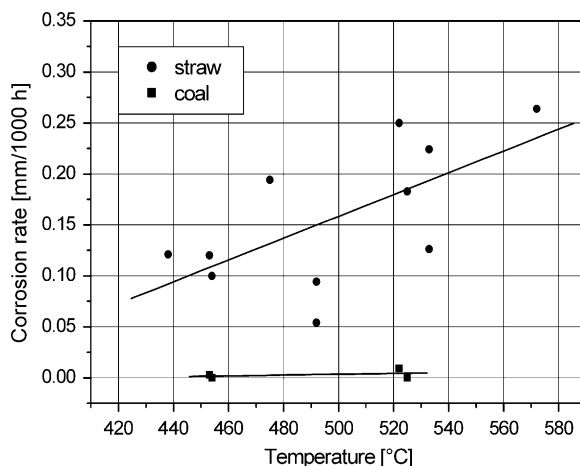
While in co-combustion of woody bio-fuels, additional operational problems are not expected, herbaceous fuels may cause severe corrosion, slagging and fouling, with the potassium chloride contained in such bio-fuels playing a major part (Binderup Hansen et al. 1997).

6.5.4.1 Co-combustion of Coal and Straw in an 88 MW_{th} CFBC

Coal and straw have been co-fired for over 10 years at the CHP plant in Grenaa, Denmark. The 88 MW_{th} circulating fluidised bed furnace is designed to fire up to 60% straw and up to 100% coal on an energy basis. During initial operation, co-firing of straw and a coal type with a high sulphur content of 3% (each with a 50% thermal input fraction) resulted in severe slagging in the furnace, in the cyclone and in the superheater area, so that the operational parameters could not be maintained even shortly after start-up. In consequence, coals with a sulphur content below 1% were later used exclusively, and deposit formation occurred only to a minor extent. The only remaining problem was in the superheater area, where the narrowly designed spacing (tube pitch) of 37 mm was favourable for the build-up of deposits. The measures used to prevent deposits in this area are to employ hanging superheaters with a spacing of 50 mm (tube pitch) (thus avoiding bridging) and to lower the flue gas temperature at the superheater. At flue gas temperatures below the melting point of potassium chloride (770°C), solid deposits do not pose any problem. In comparison to combustion of coal only, the fouling rate from using a straw fraction of 50% quintupled, though the deposits could be removed easily (Binderup Hansen et al. 1997; Clausen and Sorensen 1997). Special attention is paid to the bed inventory in order to limit the enrichment of potassium in the bed and to prevent bed agglomeration. Instead of sand with a high silica content, ash from a stoker-fired furnace with a high alumina concentration is used (Wieck-Hansen and Sander 2003).

Severe corrosion was observed at the final convective superheater, which had to be replaced after only 1 year of operation. Corrosion studies were carried out in the fluidised bed furnace by testing several different materials. Using a different

Fig. 6.61 Corrosion rate during co-combustion as a function of the steam temperature when using a 50% straw fraction in a circulating fluidised bed furnace (Binderup Hansen et al. 1997)



high-alloy steel did not result in any substantial improvement. The results, although uncertain considering the short test periods of 500–1000 h, revealed that considerable corrosion problems occurred on convective superheater surfaces with straw co-combustion in the circulating fluidised bed furnace. For martensitic steel, type X 20CrMoV 12 1, the corrosion rate was about one order of magnitude higher in straw co-combustion than in coal mono-combustion, and also considerably higher than in pulverised fuel firing with the same straw fraction (see Fig. 6.61). The cause of these high corrosion rates is assumed to lie in the in situ desulphurisation in the fluidised bed, which favours the formation of potassium chloride. The potassium chloride condenses on the superheater tubes, where it forms potassium sulphate, releasing the corrosive chlorine in the process. In contrast, pulverised fuel firing has less potassium sulphate and HCl, and hence lower corrosion rates (Henriksen et al. 1995). The principles of corrosion are discussed in detail in Sect. 5.10.4.

Various measures were implemented to reduce the rate of corrosion of the convective superheater. The main measure was the reduction of the bed temperature by about 60°C, so that it is now 860°C or lower. The changes were successful to the extent that the superheater was still in service after 7 years of operation (Wieck-Hansen and Sander 2003).

New techniques have made the superheater an in-bed heat transfer surface in the fluidised bed rather than placing it in the flue gas path. However, perhaps because there may still have been unburned straw particles in the cyclone return pipe, which form KCl as they burn, corrosion probes inserted in the return pipe measured comparably high wear rates. In order to avoid corrosion, it became necessary to arrange an uncooled section upstream of the fluidised bed superheater to ensure complete combustion. This design turned out to be successful, though erosion occurred after several years (Wieck-Hansen and Sander 2003).

Uses for the mixed ashes of this coal/straw CFBC have not yet been found, and so they have been disposed of in landfills to date (Clausen and Sorensen 1997). There was a significant reduction of N₂O emissions from the 88 MW_{th} circulating

fluidised bed furnace in comparison to coal firing alone, a fact explained by the higher temperatures in the upper part of the furnace and in the cyclones. The chlorine input into the process using straw at a thermal input fraction of 60% was 20 times higher than in the combustion of coal only. The inputted chlorine was found almost entirely in the flue gas.

6.5.4.2 Co-combustion of Sewage Sludge in a CFBC

Tests in a 230 MW_{th} brown coal-fired circulating fluidised bed for sewage sludge co-combustion have shown that emissions of SO₂, NO_x, CO and dust are within the normal operational range. The mechanically dewatered sewage sludge has a moisture content of 70%. This experience has demonstrated that the performance of circulating fluidised bed furnaces does not deteriorate through sewage sludge co-combustion. Depending on the calorific value of the sludge input, there is only a drop in the thermal output and thus in the steam production. The percentage regulation limits imposed by the 17th *BImSchV* (17th Amendment to the Federal German Pollution Control Act) were not reached. For continuous operation, an additional flue gas cleaning stage, consisting of an entrained-flow absorber using lignite-derived coke, was installed downstream of the ESP in order to ensure compliance with the mercury limits (Bierbaum et al. 1996). The plant has been in continuous operation since 1995 and co-fires approximately 200,000 t of sewage sludge a year (Roper and Kipshagen 2003).

Given that sewage sludge has an ash content of about 15% (raw), the ash load increases considerably. The pollutants in the sludge are captured inertly in the ash, except for mercury, which is transported in the flue gas. The ash is, as before, utilised for regeneration of opencast brown coal mines. It meets the requirements for land-fill grade 1 of the German Technical Instructions on Municipal Solid Waste (*TA Siedlungsabfall*).

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